ENERGY OPTIMIZATION OF THE PRODUCTION OF CELLULOSIC ETHANOL FROM SOUTHERN PINE

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ENERGY OPTIMIZATION OF THE PRODUCTION OF CELLULOSIC ETHANOL FROM SOUTHERN PINE

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LIST OF ABBREVIATIONS

EIOLCA	Economic Input Output Life Cycle Assessment
HMF	Hydroxymethylfurfural
ILEA	Institute for Lifecycle Environmental Assessment
MTBE	Methyl Tertiary Butyl Ether
MVR	Mechanical Vapor Recompression
NER	Net Energy Ratio
NREL	National Renewable Energy Laboratory
ODT	Oven Dry Ton
OPEC	Organization of Petroleum Exporting Countries
PT	Pre-Treatment
SSF	Simultaneous Saccharification and Fermentation
WWT	Waste Water Treatment

SUMMARY

On the forefront of the recent expansion in biofuels research is the production of cellulosic ethanol, or ethanol produced from a cellulose containing feedstock. Cellulose is a six-carbon polysaccharide found in most plant life and is one of the most abundant organic compounds on the planet. While the first generation of ethanol facilities uses sugar and starch based (corn kernels) plants as their feedstock, the next generation will use cellulosic sources such as wood chips, switchgrass, and forest residues. These cellulosic sources require far less energy and resources to grow and harvest, and are also much more abundant.

A cellulosic source widely available in Georgia and much of the southeastern US is southern pine. This study involves the modeling of a complete 2000 dry ton per day pine to ethanol production facility with the AspenTech[3] software Aspen Plus, which outputs a mass and energy balance as well as the capital cost of the equipment. A key parameter which affects the competitiveness of cellulosic ethanol is the internal processing energy required to convert the pine to ethanol. As a result, the heat and electrical load of every component within the facility is modeled and then quantified through the Aspen Plus simulation. After this base case energy analysis is developed, various alternate plant configurations are integrated in an attempt to reduce this process energy requirement.

The material that is not fermented into ethanol is burned on-site to provide steam and electricity to the plant, as well as excess electricity to be sold to the grid as a byproduct. As the facility processing energy requirement is decreased, more excess electricity is available for sale. The

implementation of the alternate distillation scenarios effectively reduce the internal processing energy in a manner as to increase the amount of excess electricity sold to the grid by 13.5%. The additional equipment required in this alternate scenario returns a simple payback period of 1.1 years through the additional revenue of the increased electricity sale.

The amount of net renewable energy generated per unit of non-renewable energy consumed is also increased from 10.8 to 11.2, and the amount of net life cycle carbon reductions is increased from 90.9% to 91.2%.

Chapter 1: Introduction

Throughout the past several decades there have been bursts of research in to the production of biofuels in response to a dramatic event or driving force. At the turn of the last century there was a surge in the production of biofuels in response to the creation of the automobile, which originally combusted grain alcohol, also known as ethanol. During the Second World War the Fischer-Tropsch process was utilized in Germany to produce liquid fuels during times of limited trade and high energy demand. These liquid fuels were derived primarily from coal, but research was also performed using biomass as a feedstock. Also, a resurgence of ethanol production was experienced in Germany to supplement gasoline supplies, mainly with potatoes as the feedstock. Finally, in the 1970's and early 80's there was significant research into the production of biofuels in response to the OPEC initiated oil crises in the US as well as many other parts of the world.

All of these bursts of research were in response to a single driving force; the creation of an industry with the invention of the automobile, the need for a domestic fuel during the Second World War, and again the need for a domestic fuel during the OPEC crisis. As a result of these short periods of research, progress was made in the commercialization of this industry, however once these temporary driving forces ended so did the research. These events initiated research and made significant strides, however due to their brevity did not establish the biofuels industry to stand on its own.

We are once again experiencing a resurgence in the research of biofuels, however this time it is not in response to a single driving force, but to three independent forces. These forces are the

need to develop a domestic fuel, the need to develop a renewable fuel, and the need to develop a fuel which does not add net carbon to the atmosphere. With the strength of these three independent driving forces behind it the US biofuels industry may finally be on its way to becoming a self-sustaining commercial scale industry.

Leading the recent growth of the US biofuel industry is the production of ethanol from corn kernels, with an annual capacity anticipated to exceed 7 billion gallons in 2007, over 40% greater than in 2006. This dramatic growth is in response to the continuation of the federal subsidy (ethanol blenders credit), the implementation and expansion of the Renewable Fuels Standard, rising crude oil prices, the phase out of MTBE (methyl tertiary-butyl ether) as an oxygenate for gasoline blending, as well as a less quantifiable recent nationwide shift to support a domestically produced fuel with a smaller carbon footprint. However, along with these increased production levels has come an accompanying increase in corn prices which has recently decreased the profitability of these facilities. Furthermore, as an increasingly large portion of the US corn market has become devoted to ethanol production, a strain has been placed on the nation's corn-based food supply. As a result, to continue the growth of this industry, a diversified feedstock base must be developed which extends beyond the use of corn kernels.

On the forefront of this biofuels research is the production of cellulosic ethanol, or ethanol produced from a cellulose containing feedstock. Cellulose is a six-carbon polysaccharide found in most plant life and is one of the most abundant organic compounds on the planet. While the first generation of ethanol facilities uses sugar and starch based (corn kernels) plants as their feedstock, the next generation will use cellulosic sources such as wood chips, switchgrass, and

forest residues. These cellulosic sources require far less energy and resources to grow and harvest and are also much more abundant. The US department of agriculture recently surveyed the amount of biomass available in the US which could be sustainably used by the biofuels industry and set this value at over one billion dry tons per year[4], with the vast majority of this biomass being cellulosic in nature. This large cellulosic feedstock resource puts the US in the position to produce approximately 30% of its transportation fuel domestically, renewably, and with a decreased carbon footprint.

A cellulosic source widely available in Georgia and much of the southeastern US, and investigated in this study, is southern pine. Southern pine has been grown in Georgia for well over a century and as a result a strong and established infrastructure exists for its planting, managing, harvesting, and distribution. The majority of the pine demand has traditionally come from the pulp and paper industry. However, due to recent increases in foreign competition as well as increased environmental standards placed on these facilities, the industry has downsized in the southeastern US. This decreased demand for pine, with a now overly-developed infrastructure, presents an opportunity for the large scale deployment of ethanol producing facilities with southern pine as the feedstock throughout the southeastern US.

With this developed infrastructure as well as a deep understanding of the growth of southern pine, the southeastern US has the ability to produce this feedstock in an efficient manner, dramatically decreasing the energy demand associated with the production of biofuel feedstocks. For traditional feedstocks, such as corn kernels, much of the benefit of producing renewable ethanol is offset by the use of non-renewable fuel in the agricultural phase. When using southern

pine as a feedstock, much of the benefits of producing the ethanol are preserved. A metric which measures this trade-off is called the net energy ratio (NER) and is discussed further in this study.

The reduced agricultural energy in the production of pine versus corn also decreases the carbon footprint of producing ethanol from southern pine. The carbon held within the pine is taken exclusively from the atmosphere in the form of carbon dioxide. When the tree matter is processed into ethanol and combusted, the only carbon released is the carbon dioxide that was originally extracted from the atmosphere resulting in a closed loop system. The only net carbon released during the process is from the combustion of fossil fuels required in the agricultural phase to grow the pine as well as the transportation of the inputs and products. The result is a severe reduction in the amount of net life cycle carbon released into the atmosphere as opposed to the production of ethanol from corn kernels or from the production and combustion of gasoline.

Overall, cellulosic ethanol produced from southern pine has the potential to be produced domestically at a large scale, in a dramatically more renewable fashion, and with a much smaller carbon footprint. These impacts are quantified and discussed in the following sections of this study.

Chapter 2: Literature Review

The state of the technology of the ethanol production industry was reviewed prior to performing this study. Several sources were reviewed which dealt with the overall topic of producing cellulosic ethanol, as well as sources that focused on one or more topics such as ethanol fermentation, pre-treatment, ethanol-water distillation, and economic modeling. Further studies investigated the sensitivities of parameters to the minimum required selling price of ethanol such as the fermentation solids loading, yeast and enzyme concentration, pre-treatment conditions, and others. All of these studies were used as guides to develop an understanding of the industry prior to performing this research.

National Renewable Energy Laboratory - 2002[1]

One of the key recent analyses which investigates the production of transportation grade ethanol from cellulosic sources was developed by the National Renewable Energy Laboratory (NREL) and published in 2002. This study models the use of corn stover (cellulosic source) as a feedstock and produces ethanol through dilute acid hydrolysis and simultaneous saccharification and co-fermentation. This extensive study used NREL performed experiments to calculate yields of different sections of the process. The entire process was then rigorously modeled with the software Aspen Plus[3], incorporating the experimental yields where necessary. This study also performed an economic analysis by developing an installed cost for each component within the plant along with all fixed and operating costs needed for such a facility. The end result was a minimum required ethanol selling price, which was shown to be very cost-competitive with gasoline.

Many of the assumptions and correlations developed in this NREL study were used to create the base case model presented in this study. Also, it was the work performed during this NREL study that made it possible to bring the non-traditional cellulosic biomass database into the Aspen Plus simulation component database. This study was a valuable resource, however it made no attempt to study advanced heat integration and no sensitivities were performed with various distillation and solids recovery scenarios.

Results – This NREL study provided the foundation for creating the base case model developed in this study and also allowed the incorporation of cellulosic materials into the Aspen Plus simulation environment. Many of the assumptions and correlations for the peripheral plant components were taken from this NREL study. This NREL study itself concluded that ethanol derived from corn stover can be produced in a cost-competitive manner to gasoline with today's technologies, and in an even more competitive fashion in the future.

National Renewable Energy Laboratory - 1999[23]

Another key reference analyzing the production of cellulosic ethanol is the 1999 NREL report (NREL 1999), the pre-cursor to the 2002 NREL report discussed above. The primary difference between the reports is that this 1999 study used yellow poplar instead of corn stover as its cellulosic source. Virtually all of the assumptions, component models, and results were updated in the 2002 report and therefore are more relevant and accurate. However, the fact that this 1999 study used a woody biomass as an input gave insight into the feedstock handling section developed in this study. It also provided beneficial insight into the differences between using a

woody biomass versus corn stover as a feedstock, such as differences in input moisture content and how this affects the overall plant operation. This study provided great insight, however no attempt was made to vary distillation and solids recovery scenarios as is the focus of this study.

Results – This 1999 NREL report gave insight as to the handling and processing of a woody biomass as opposed to corn stover in the 2002 NREL report. This NREL report concluded that the ethanol produced from the yellow poplar was cost-competitive with gasoline, however not as much as corn stover as the NREL 2002 report concluded three years later.

Lund University

Several key studies were reviewed from the Department of Chemical Engineering at Lund University in Sweden. These studies primarily used the cellulosic feedstock of spruce, which is similar to southern pine, and therefore a strong relationship should exist between the results obtained by this research and comparable work to be performed on pine. These studies analyzed the:

- differences between separate hydrolysis and fermentation (SHF) and simultaneous saccharification and fermentation (SSF)[19]
- sensitivities of pre-treatment yields to acid concentration, pre-treatment temperature, and residence time[16]
- effect of chip washing in one and two step pre-treatment scenarios[15]
- use of sulfuric acid as opposed to sulfur dioxide[14]

They also performed several plant-wide economic analyses for the economical production of cellulosic ethanol[18, 20, 21]. All of these articles were reviewed to understand the state of the science and also to help build the base case for this current study.

Results – These Lund University studies provided the information necessary to develop the yields and process flow within the saccharification and fermentation sections of the facility. The reports themselves concluded that ethanol derived from softwood is a viable process, and that spruce is an attractive feedstock resulting in high yields. The process and economic models also conclude that this process can be cost-competitive to gasoline when produced at commercial scale.

Wingren 2005[22]

A study was performed by Anders Wingren at Lund University specifically looking at energy considerations in the production of cellulosic ethanol. This research involved producing ethanol from spruce in a process using sulfur dioxide impregnation and SSF, similar to what is considered in the base case presented in this study. The energy optimization scenarios considered were to increase the number of effects in the evaporator, integrate a stripper with the evaporator, use mechanical vapor recompression within the evaporator, and the use of an anaerobic digestor in place of the evaporator. All of these scenarios were modeled and the results were tabulated including total heat requirement, total capital cost, variable costs, and fixed costs. From these results the minimum ethanol selling price for each scenario was presented and discussed.

Wingren's study is similar to the analysis performed in this study, however it uses a feedstock of spruce instead of pine. Also, the pre-treatment and SSF conditions and yields are different, resulting in dramatically different conditions at the inlet of the distillation section (total mass flow, ethanol mass flow, ethanol concentration, solids mass fraction, etc.). The SSF solids loading is lower in Wingren's study (8.4% vs. 10%), which dramatically changes the amount of energy required in the distillation and solids recovery section. The steam generation from burning the solids residue is also different in Wingren's study as compared to this study, since only enough steam is generated to supply all heat duties on-site. The remaining solids are pelletized and sold as byproduct. Generating all steam and electricity on-site can help to better quantify the costs of using varying qualities of steam from the extraction points of the power generation turbine. Finally, the actual heat integration scenarios described above used in Wingren's study vary from the scenarios analyzed in this study.

Results – Wingren's study provided a sample comparative analysis of various energy reducing scenarios, however due to different feedstock and processing parameters direct performance correlations can not be drawn between Wingren's study and this study. Wingren's study concludes that there is significant cost and energy saving potential in the exploration of heat integration scenarios in softwood to ethanol production facilities

Comparative Net Energy and Carbon Assessment – Berkeley[8]

This Berkeley study evaluated the results presented in six published reports by different authors as to the net energy and net life cycle green house gas emissions associated with the production of ethanol from corn and other feedstocks. This Berkeley study then attempted to make

adaptations to these six independent analyses to equalize the assumptions and present the results on a level field. General cases for current corn ethanol production, carbon dioxide intensive corn ethanol production, and cellulosic ethanol production were then developed by this research group.

This article concluded that, with the exception of one outlier, the production of corn ethanol has a net energy gain as well as a net reduction in carbon emissions. Furthermore, it concluded that the production of ethanol from cellulosic material had a dramatically higher net energy gain and higher net reduction in carbon emissions.

Results – The analysis is an example of how to perform a rigorous comparative energy and carbon life cycle assessment, and shows that the production of ethanol from corn and especially from cellulose is an improvement over gasoline production from fossil fuels. This Berkeley study concludes that although there is some variation among analyses, as well as an occasional outlier, the net energy and net life cycle carbon emissions of the production of corn ethanol is consistently favorable compared to gasoline.

Comparative Net Energy and Carbon Assessment – ILEA[11]

This ILEA (Institute for Lifecycle Environmental Assessment) study also evaluates several published reports which compute the net energy of the production of ethanol from corn and cellulosic sources. Differences between the assumptions amongst the different studies are noted, however no attempt is made to numerically correct and normalize these assumptions. It is shown that of the six reports which evaluate the production of ethanol from corn, five show a net energy

gain ranging from 1.29 to 1.65 (ratio of renewable output energy to non-renewable energy input). The sixth report shows a net energy loss (0.84), however this value is still less of a loss than incurred in the production and distribution of gasoline (0.76)[9].

This ILEA study also examines the production of ethanol from cellulosic sources presented in four reports. These studies all analyze different cellulosic sources, and the range of the net energy values for three of the articles is from 4.40 to 6.61. The fourth paper, the same as which shows a net energy loss for corn ethanol, shows a net energy ratio of 0.69 for the production of ethanol from switchgrass. This fourth report is labeled an outlier and not considered credible.

Results – This paper is another example of how to perform a comparative net energy analysis, and shows that the production of ethanol from corn and cellulosic sources exhibit a net energy gain.

Chapter 3: Project Scope

There are two main obstacles impeding the further growth of ethanol production in the US. The first is the cost competitiveness of ethanol with that of gasoline. Although the selling price of ethanol is currently higher than its production cost, this selling price is related to current gasoline prices. Therefore, if crude oil, and in turn gasoline prices, were to decrease the selling price of ethanol would also decrease, possibly to a situation which would make currently produced ethanol unprofitable. Adding to this issue is the fact that ethanol is currently heavily subsidized through federal and state incentives. If these subsidies were to lapse the cost-competitiveness of ethanol would be in further jeopardy.

The second obstacle is the issue of the net energy analysis. As mentioned previously, the ratio of the amount of non-renewable input energy to the renewable energy output is referred to as the net energy ratio (NER), and is an important metric in the biofuels industry. Current ethanol production from corn has a range of NER's generally between 1.0 and 1.5[8, 11]. This is to say that for every unit of energy invested to produce ethanol from corn, 1.0 to 1.5 units of energy are returned in the heating value of ethanol along with the energy value of its co-products. The NER of cellulosic ethanol varies widely depending on the process and the feedstock. The NER of the cellulosic ethanol produced in the process presented in this study using the feedstock of southern pine has been calculated and is shown to be several times higher than corn ethanol.

A process improvement which would help to alleviate both of these impediments is to decrease the processing energy used in the production of ethanol. In the process modeled in this study,

the solid residue remaining after fermentation is combusted on-site to create steam and electricity for the plant, as well as excess electricity to be sold to the power grid as a byproduct. By decreasing the steam and electrical demand within the plant more electricity is sold to the grid. This increased sale of electricity increases the cost-competitiveness of ethanol by increasing a revenue stream, and also increases the process NER by increasing the energy of the byproduct for the fixed energy inputs.

To quantify the heat and electrical loads the entire plant has been modeled using the AspenTech[3] software Aspen Plus 2004.1. This modeling software has the thermodynamic and chemical capabilities to capture the dynamics of all of the processes within the plant. A base case was first developed by integrating information from several sources and adapting them to use southern pine as the feedstock. This base case process was rigorously modeled and then the results were compiled and a mass and energy balance performed. With this information all heat and electrical loads can be quantified and allocated to their specific component and section.

Several different scenarios are considered in an effort to reduce the energy consumption with respect to this base case. Energy is used in virtually every component within the plant, however, the vast majority (~2/3) of the heat and electrical loads are in the distillation and solids recovery section. As a result, only the distillation and solids recovery section will be altered in these alternative scenarios. All other sections of the plant will remain fixed at the base case conditions when these other scenarios are modeled.

The base case and all of the alternative scenarios will be ranked for effectiveness, with the two overall figures of merit being the amount of excess electricity available for sale and the highest NER. The change in incremental capital for each of these scenarios is then evaluated and the simple payback period for each is calculated.

Chapter 4: Modeling and Methodology

To rigorously obtain the heat and electrical loads, as well as material flow rates, temperatures, pressures, and compositions the entire processing plant was modeled with the simulation software Aspen Plus 2004.1. A base case model was first developed to set a standard for all other scenarios to be compared against. All of the assumptions made in this base case model are based off of the reviewed literature and internally performed experiments.

The plant is broken down into nine sections; feedstock handling, pre-treatment, SSF, yeast addition, distillation and solids recovery, waste water treatment, storage, power generation, and utilities. The primary process flow of these sections is shown below in Figure 1, and then each of the sections are discussed individually in detail.

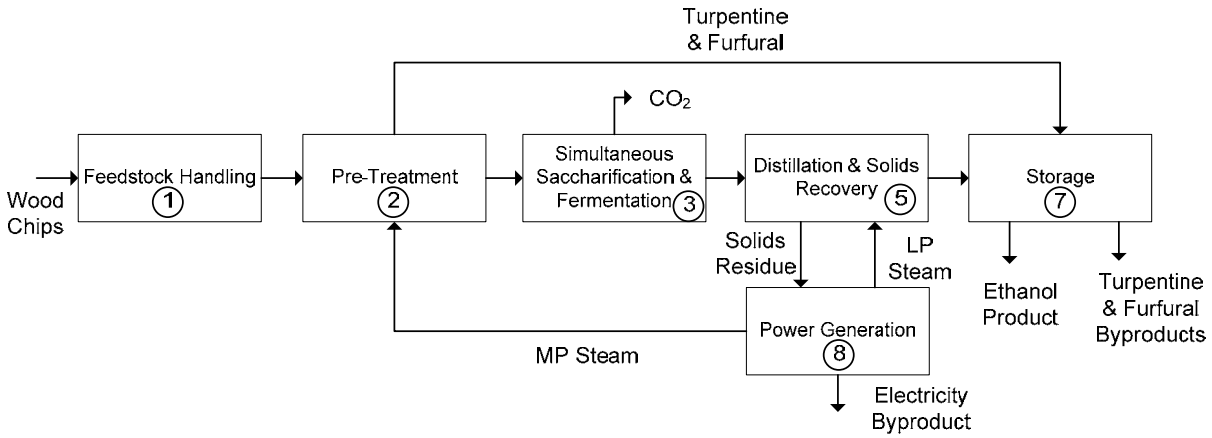

Figure 1: Primary process flow diagram

4.1 Section 1 - Feedstock Handling

Methodology

The feedstock handling section entails receiving the wood chips from the delivery trucks and preparing them to enter the plant. This includes washing the chips to remove dirt and debris as well as re-chipping any wood chips which arrive at the plant oversized. This section ends where the cleaned and chipped wood chips enter the conveyor belt to the pre-treatment section.

Process Description

The base case plant consumes 2000 dry tons (4000 wet or green tons) of pine chips per day which it receives by truck. The plant receives approximately 160 trucks per day, each carrying 25 green tons of chips, and each of the two truck dumping/chip stacking systems is sized so that one truck can unload in 15 minutes.

The trucks are weighed, loaded, and unloaded on a combination scale and hydraulic dumper that dumps the load into a receiving hopper. A vibrating conveyor meters the chips onto a transfer belt conveyor, which transports them to a radial stacking conveyor. The stacking conveyor deposits the chips onto a large chip pile that holds approximately seven days worth of chips. A front end loader reclaims chips by scooping and dumping the chips into a reclaim hopper. The reclaim hopper feeder meters the chips onto the reclaim hopper conveyor that transports the chips to the washer surge bin. A magnet separator removes tramp metal while the chips are on the reclaim hopper conveyor.

The chip wash feeder meters the chips out of the washer surge bin into a turnkey chip washing system. The scalper screen feeder transports the chips from the washing system to the scalping screen, which removes grossly oversized chips. Chips then proceed to the chip thickness screen. Approximately 20% of the chips are considered oversized and are reduced in size in the disc refiner. Acceptable chips are transported to the pre-treatment section on the pretreatment feeder.

4.2 Section 2 – Pre-Treatment

Methodology

The pre-treatment section uses thermodynamic (pressure and heat), mechanical (grinding), and chemical catalytic (acid soaking) processes to begin to breakdown the complex reinforced structure of the wood. The southern pine chips used in this process consist primarily of cellulose, hemi-cellulose, and lignin. The complete composition of southern pine modeled in this study is shown below in Table 1.

Table 1: Composition of Southern Pine

Component	% Dry Basis	% Wet Basis	Inlet Mass Flow (lbs/hr)
Glucan (Cellulose)	42.7	21.4	71,167
Xylan	6.0	3.0	10,000
Galactan	2.5	1.3	4,167
Arabinan	1.1	0.6	1,833
Mannan	12.9	6.5	21,500
Others	0.4	0.2	667
Lignin	31.2	15.6	52,000
Ash	0.4	0.2	667
Protein	0.4	0.2	667
Extractives	2.4	1.2	4,042
Moisture	-NA-	50.0	166,708
			333,417

Cellulose is a 6-carbon carbohydrate polymer which makes up the bulk of the wood and tends to be in the geometric center of the plant. The lignin is a reinforcing structure which surrounds the cellulose, and the hemi-cellulose is a combination of 5 and 6-carbon polymers (mannan and galactan are 6-carbon and xylan and arabinan are 5-carbon polymers) which fills in the gaps. The job of the pre-treatment section is to break apart this structure to reveal the individual components of the wood and leave them vulnerable to further hydrolyzation by enzymes and yeasts in the SSF section.

The conditions within pre-treatment (temperature, acid concentration, residence time) must be finely tuned for each feedstock, because over-treating of the wood will degrade the carbohydrates and leave them unable to be fermented. Any degraded forms of 5 and 6-carbon sugars, furfural and hydroxymethylfurfural (HMF) respectively, also act as inhibitors in the fermentation process. This can result in longer required fermentation residence times as well as lower fermentation yields. The pre-treatment conditions for this base case are a temperature of 215 C (419 F), a residence time of five minutes, and a sulfur dioxide acid concentration of 2 %w/w of the incoming water content. Through a review of the current literature as well as internally performed experiments, the reactions occurring in this process have been calculated and are shown below in Table 2.

Table 2: Pre-Treatment Reactions and Yields

Stoichiometric Reaction	Fractional Conversion
H2O + 2 CELLULOS(Cisolid) --> CELLOBIOSE	0.001
H2O + CELLULOS(Cisolid) --> GLUCOSE	0.080
CELLULOS(Cisolid) --> HMF + 2 H2O	0.001
XYLAN(Cisolid) --> XYLOLIG	0.1
H2O + XYLAN(Cisolid) --> XYLOSE	0.600
XYLAN(Cisolid) --> FURFURAL + 2 H2O	0.150

MANNAN(Cisolid) --> MANOLIG	0.1
H2O + MANNAN(Cisolid) --> MANNOSE	0.600
MANNAN(Cisolid) --> HMF + 2 H2O	0.150
GALACTAN(Cisolid) --> GALAOLIG	0.1
H2O + GALACTAN(Cisolid) --> GALACTOSE	0.600
GALACTAN(Cisolid) --> HMF + 2 H2O	0.150
ARABINAN(Cisolid) --> ARABOLIG	0.1
H2O + ARABINAN(Cisolid) --> ARABINOSE	0.600
ARABINAN(Cisolid) --> 2 H2O + FURFURAL	0.150
LIGNIN(Cisolid) --> LGNSOL	0.08

In the above table, the components labeled as Cisolid indicate that these are part of a separate solid substream within the liquid stream. The pre-treatment process in general hydrolyzes (breaks down by the addition of water) these solid polymers into soluble simple sugars (glucose, mannose, galactose, xylose, and arabinose). The compound cellubiose is simply two glucose molecules joined together, and occurs when a cellulose carbohydrate is not fully hydrolyzed into separate glucose molecules.

In the case of overly harsh pre-treatment material can also be degraded to oligomers, or a compound similar to lignin (pseudo-lignin). These compounds are shown in the table above as xylolig, manolig, galaolig, and arabolig. These compounds are no longer able to be converted to ethanol and simply act as bystanders throughout the process until they are burned in the power generation section.

A variety of compounds are produced when the chips are combined with acid at these high temperatures including turpentine, the inhibitors furfural and hydroxymethylfurfural, disassociated liquid sulfurous acid (H_2SO_3), and gaseous sulfur dioxide. All of these chemicals

are harmful to the fermentation process and an attempt to remove them from the process stream is made through a series of flashes. These compounds are drawn off in order of highest volatility as the pressure of the flow drops from 350 psi to atmospheric. The compounds can then be sent back to storage tanks to be re-used in the process (SO_2) or further separated and stored as valuable by-products (turpentine and furfural).

Process Description

When pine chips enter the pre-treatment section from the chip conveyor coming from feedstock handling they are first mixed with gaseous sulfur dioxide (SO_2) in a pre-soak vessel. The chips are soaked in SO_2 for approximately 20 minutes and then are moved to the pre-treatment vessel by a screw conveyor. In the first stage of the pre-treatment vessel 65 psi steam is first injected to raise the temperature of the medium to 100 C, live steam at 350 psi is then injected in the main chamber to bring the temperature of the mixture to 215 C (419 F). The chips are pre-treated for 5 minutes and are then flashed into the first blowdown tank at 1 atm.

This first blowdown tank is jacketed and maintained at atmospheric pressure and 100 C, resulting in approximately 90% of the SO_2, 75% of the furfural, and 60% of the water to flash off as vapor. This vapor proceeds to a second tank which is jacketed and maintained at 1 atm and 75 C. In this vessel, approximately 95% of the entering SO_2, 10% of the furfural, and 5% of the water leaves as vapor and is recycled to the pre-treatment pre-soak vessel. The liquid from the second blowdown tank consists of water, furfural, and turpentine. These three liquids have significantly different densities and are separated by a decanter. The turpentine and furfural are

sold as by-products and the remaining water is sent to the waste water treatment section for purification.

The bottoms from the first blowdown tank are then sent to the SSF section through a screw conveyor.

4.3 Section 3 – SSF (Simultaneous Saccharification and Fermentation)

Methodology

The SSF section first uses enzymes to hydrolyze any remaining carbohydrate solids into simple sugars. These sugars, along with sugars released during pre-treatment, are then metabolized by yeast into ethanol and carbon dioxide. The enzymes and yeasts used in this process are of considerable cost and are therefore used sparingly, even though an increased concentration of either will generally increase ethanol yield and decrease SSF residence time. It is an optimization step to calculate for a specific process the concentration of yeasts and enzymes, as well as residence time and operating temperature, which will be the most cost-effective. For this base case plant these parameters have been set to 2 grams of yeast per liter of SSF material, 15 FPU (filter paper units) of enzymes for every gram of cellulose, 72 hours, and 37 C.

Another high-influence parameter in the SSF section is the level of solids loading. The yeast cells work best in low concentrations of solids and ethanol, and high concentrations of water. Therefore, higher yields can be achieved with a lower solids loadings, which is defined as the mass of water insoluble solids (WIS) divided by the total mass. Recycle water is added from

throughout the plant to the fermentation vessels to achieve this level of solids loading, however while a decreased solids loading increases ethanol yield it also dramatically increases the energy required to purify the ethanol in the distillation section. This again is an optimization problem, to trade-off the increased yields versus increased steam and electrical demand. For this study a solids loading of 10% is used, which is considered at the lower end of the spectrum (more water, less solids) as far as possible loadings. From a review of the literature as well as internally performed experiments the SSF yields have been calculated for this base case and are shown below in Table 3.

Table 3: SSF Reactions and Yields

SSF Reaction	Fractional Conversion
GALACTOS --> 2 ETHANOL + 2 CO2	0.92
MANNOSE --> 2 ETHANOL + 2 CO2	0.92
GLUCOSE --> 2 ETHANOL + 2 CO2	0.92
GALACTAN(Cisolid) + H2O --> 2 ETHANOL + 2 CO2	0.81
MANNAN(Cisolid) + H2O --> 2 ETHANOL + 2 CO2	0.81
CELLULOS(Cisolid) + H2O --> 2 ETHANOL + 2 CO2	0.83

It is noted that only the six carbon carbohydrates and sugars are fermented in this process. This is a result of the fermentation organism selected, Saccharomyces cerevisiae, which does not metabolize five carbon sugars. This organism, also known as common compressed bakers yeast, is selected because of its widespread availability, low cost, and proven history fermenting sugars. It is also noted that even among cellulosic sources southern pine has a relatively low mass fraction of five carbon sugars; 7.1% on a dry basis. As a result, not a large percentage of the input mass is non-fermentable because of the selection of compressed bakers yeast as the fermenting organism.

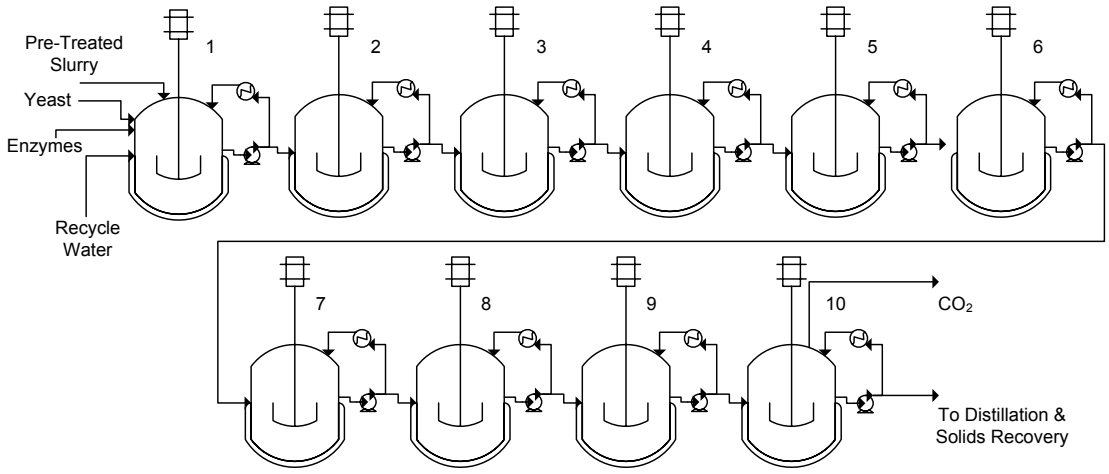

Figure 2: Ten module SSF section Process Flow Diagram

Process Description

The flow from the bottoms of the first flash vessel in the pre-treatment section is sent to the first SSF vessel. There are 10 SSF vessels in total placed in series, all of which are agitated, maintained at approximately 37 C through the use of pump around heat exchangers, and have a capacity of 1,000,000 gallons. Yeast, enzymes, and recycled water are also added to this first SSF vessel to achieve the settings discussed above.

The mixture is constantly transferred to the next vessel by a series of pumps and remains in each vessel for approximately 7.2 hours. After the tenth vessel the flow passes through a CO_2 vent where the vapor exits and passes through a venturi scrubber. In this scrubber the vapor is washed with water to remove any remaining traces of ethanol, and then this near pure stream of CO_2 is vented to the atmosphere. This release of CO_2 does not contribute to the net release of carbon, since the only carbon released in this stream is directly from the pine, and the only carbon entering the pine was directly from CO_2 extracted from the atmosphere. This CO_2 stream

can have alternate paths such as a feedstock for the growth of algae or it can be compressed and sold to the beverage industry.

After leaving the tenth SSF vessel the liquid flow is first raised to 95 C by passing through a heat exchanger with the second pre-treatment flash vessel, and then to 100 C from a heat exchanger with the bottoms exiting the beer column. This flow is then pumped to the distillation and solids recovery section.

4.4 Section 4 – Yeast Growth

Yeast is not grown on-site in this base case plant. This process was reviewed in the literature[2,13], however in this study all yeast is bought from a distributor and simply stored on-site and added to the first fermentation vessel as needed.

4.5 Section 5 – Distillation and Solids Recovery

Methodology

The ethanol exiting the SSF section is in a very dilute form (~2-5 %w/w), however to be considered transportation grade ethanol and to be able to be mixed with gasoline it must be at least 99.5 %w/w pure, with the remaining half percent being water. Ethanol is strongly hydrophilic (soluble within and attached to water through hydrogen bonds due to their respective polar geometries) and requires a great deal of energy to separate. Furthermore, the amount of

energy required increases exponentially with respect to decreasing initial concentration in very dilute mixtures.

To demonstrate this phenomenon the base case model developed in this study was fixed in all parameters except for the solids loading in SSF. As this value decreases, directly decreasing the initial ethanol concentration entering distillation, it can be seen that the required separation energy increases exponential. This is shown below in Figure 3. It is noted that as the solids loading decreases the ethanol yield will slightly increase, however this relationship is not well quantified and as a result ethanol yield is held constant in the figure.

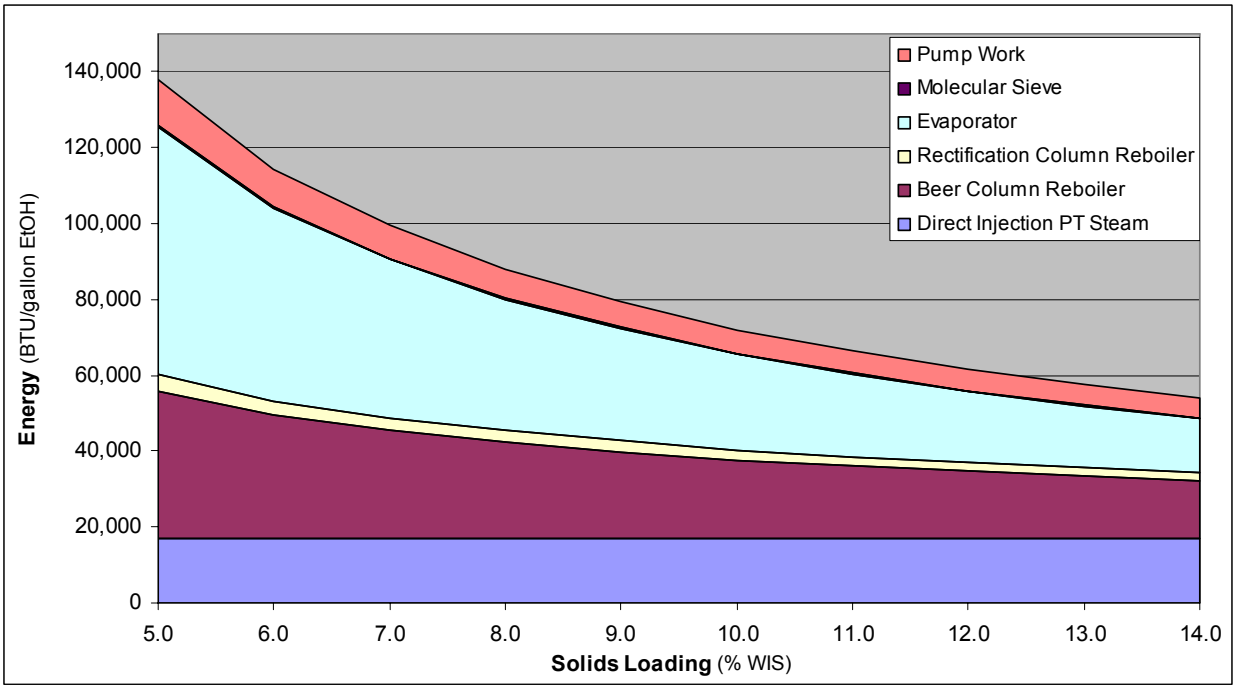

Figure 3: Plant Energy versus SSF Solids Loading

In the above figure the energy within direct inject pre-treatment (PT) steam is shown as the difference in the enthalpy of the live steam with respect to water at 1 atm and 20 C. The energy

of the beer column, rectification column, and molecular sieve is supplied by 65 psi and the evaporator energy is supplied by 25 psi steam. The pump energy shown is simply the electrical energy demand, with no correction back to primary heat energy. The valuing of these different qualities and forms of energy are discussed in Results, Section 7.0.

The primary method for separating ethanol and water is though the use of distillation, which takes advantage of the difference in volatilities to separate the compounds. The design of these distillation columns encompasses several operating parameters such as number of trays, tray height, tray efficiency, reflux ratio, and reboiler ratio. These columns can also be placed in parallel or series to collectively reach a desired purity. As a result, much attention is giving to the internal workings and parameters of each of these columns.

Ethanol and water form an azeotropic mixture at a concentration of ~94 % w/w ethanol. An azeotropic mixture occurs when the composition of the vapor resulting from boiling the mixture is the same as the composition of the base mixture itself. In this scenario, the mixture can not be further purified through conventional distillation alone. Since the required purity of ethanol (99.5%w/w) is greater than the azeotropic purity alternative methods must be employed to carry out the final stages of separation. These alternatives generally use a tertiary component to break the azeotrope, pressure swing distillation to move the azeotrope, or a molecular sieve to remove the remaining water through adsorption.

Two figures which are often used to show an azeotrope are the Txy diagram and the XY diagram. The Txy diagram shows the mass fractions of the vapor (y) and the liquid (x) phases of

the mixture as a function of temperature and mass fraction ethanol, shown as Figure 4. An azeotrope is detected whenever the vapor and liquid composition lines cross, which occurs at the ~0.95 mass fraction ethanol location at atmospheric pressure. The XY diagram, shown as Figure 5, shows the phase equilibrium line of the vapor and liquid phases overlayed against the line y=x. On this diagram, an azeotrope is detected if the equilibrium line crosses the line y=x, which occurs at approximately 0.95 mass fraction ethanol (liquid and vapor) at atmospheric pressure.

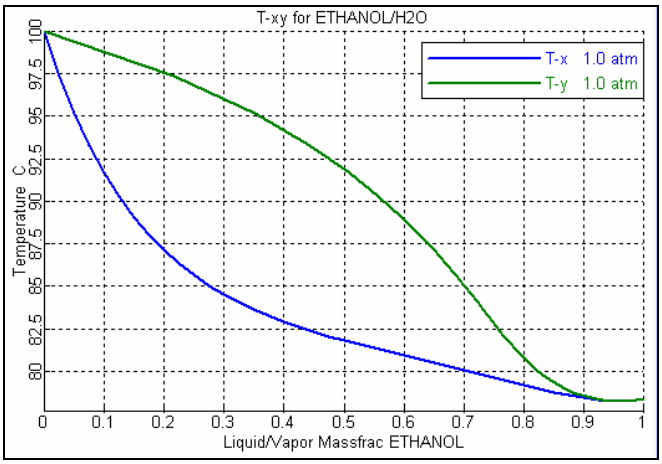

Figure 4: Ethanol Water Txy Diagram

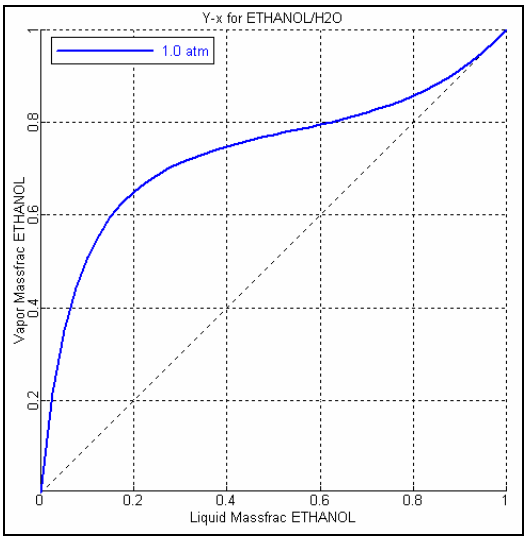

Figure 5: Ethanol Water XY Diagram

Along with purifying the ethanol this distillation section is also responsible for concentrating the solids residue to a level able to be combusted. All of the lignin, protein, and other un-fermented solids are in a dilute (2-4%w/w) mixture with water. In order to be combusted, this residue has to be dried to approximately 50% moisture. Some of the possible equipment to achieve this are air presses, multiple effect evaporators, and centrifuges.

A process flow diagram of the configuration used in this base case is shown below in Figure 6 to better describe the section layout.

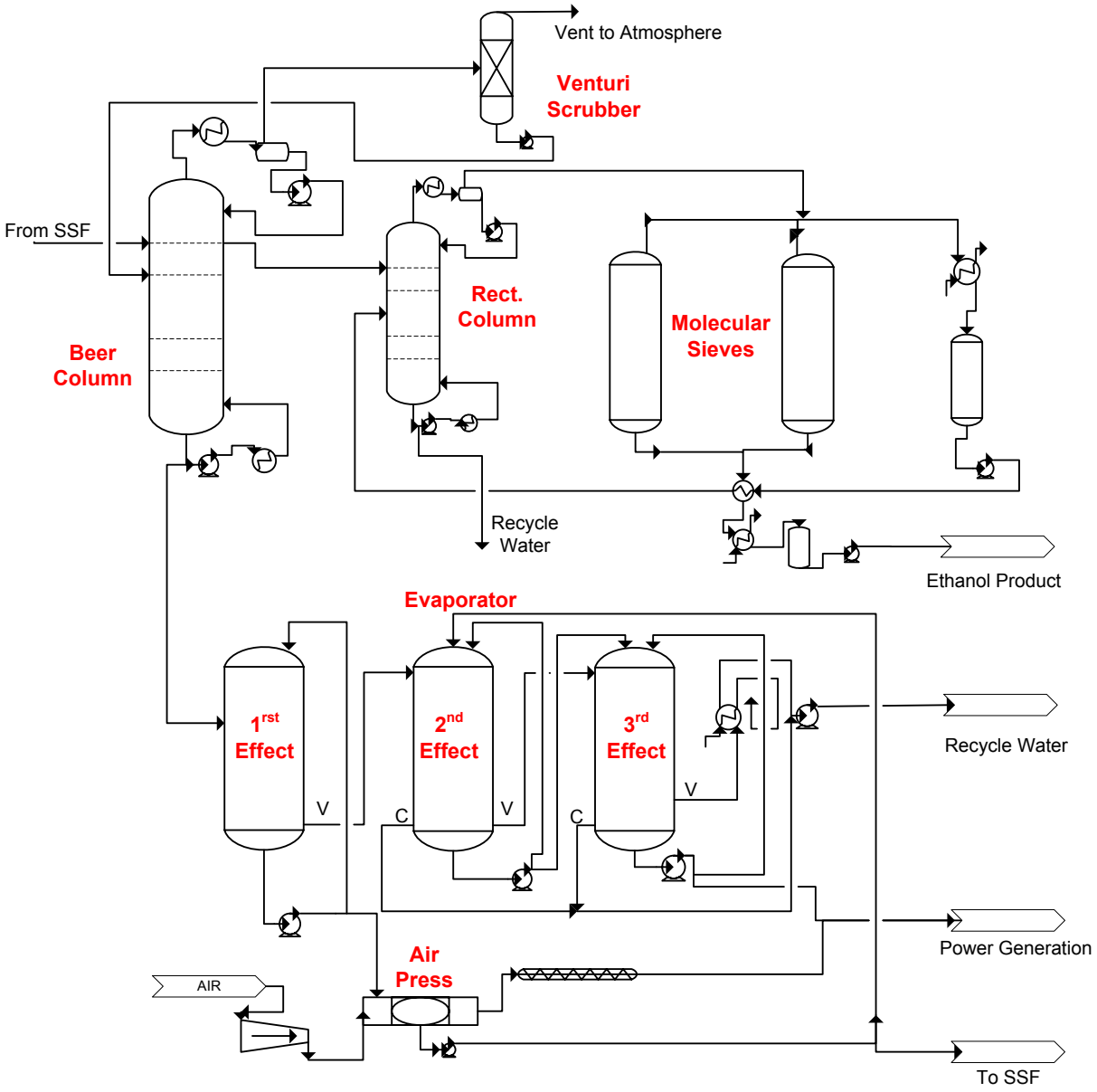

Figure 6: Distillation and Solids Recovery section Process Flow Diagram

Process Description

The dilute ethanol flow leaving SSF enters the distillation section and is pumped to the beer column (first distillation placed in series) at approximately 100 C. All of the remaining solids and a large portion of the water exit the bottoms of the beer column and the remaining CO_2 exits as a vapor distillate and is sent to a venturi scrubber. The beer column is 16 stages and has a fixed molar reflux ratio of 3.0. The dilute ethanol (3.58 %w/w) inlet enters at the second stage and the ethanol product is drawn off as a vapor at the third stage at a concentration of approximately 29 % w/w. The distillate and ethanol side draw mass flow rates are varied to achieve an ethanol mass concentration in the exiting bottoms of 0.0005 and the temperature of the condenser is set at 60 C to make sure all of the returning flow is in the liquid phase (most volatile component is ethanol, has a saturation temperature of 78 C at atmospheric pressure). The column has a slight pressure gradient with 2.10 atm at the reboiler and 1.86 atm at the condenser. Over 99% of the entering ethanol exits in the vapor side draw, over 99% of the CO_2 in the vapor distillate, and over 99% of the solids exit in the bottoms.

The ethanol in the vapor side draw exiting the beer column, along with the dilute ethanol mixture washed from the venturi scrubber, is then sent to the rectification column (second distillation column in series) for further purification. The rectification column consists of 30 trays and the ethanol exits as a distillate vapor while the majority of the water exits the bottoms as a liquid. The mass flow rate of the distillate and bottoms are varied to set the mass fraction of ethanol leaving the bottoms to 0.0005 and the mass fraction of ethanol in the distillate to 0.926. Since the top stage of this column is at an elevated temperature (~94 C), the reflux is condensed

through a heat exchanger with the first effect evaporator (~83 C) instead of cooling water, thus using some degree of heat integration to decrease the overall distillation steam demand.

The ethanol vapor distillate leaving the rectification column is sent to one of two parallel molecular sieves for final dehydration. The molecular sieves are beds packed with ~3 angstrom zeolites which have pores large enough to capture water however too small to allow ethanol molecules to enter. While one bed is operating to dehydrate the ethanol, the other bed is being recharged by having a small flow of pure ethanol run through it to extract the water from the zeolites while under a partial vacuum (~0.14 atm).

The solids which exited the bottoms of the beer column are sent to a three stage evaporator. The dilute solids flow is first throttled to a partial vacuum pressure of 0.60 atm (from 2.10 atm) and then enters the first stage of the evaporator. External heat is added to boil off much of the water which exits as a vapor and the liquid/solids flow is sent to an air press. The press uses compressed air to force the mixture through a filter which only allows liquids and soluble solids to permeate. This leaves a dried cake (~45% moisture) of insoluble solids which is sent to the power generation section, and a liquid which is sent to the second stage evaporator after being throttled to ~0.30 atm. The water vapor which exited the first stage is condensed around the second stage to provide heat which is used to boil off further water.

The bottoms of the second stage are sent to the third stage, while the exiting water vapor is condensed to provide heat to the third stage. The third stage is throttled to ~0.21 atm and sends

its exiting water vapor to a cooling water condenser and its bottoms (consisting of soluble solids at ~60% moisture) are sent to the power generation section for combustion.

4.6 Section 6 – Waste Water Treatment

Methodology

The waste water treatment section enables much of the water exiting one section to be purified and recycled to another section. If this were not done, the plant would require a large amount of make-up water and would discharge an equally large amount of waste water. The treatment consists of one or more digestors in series which consume the soluble solids, insoluble solids, and any other compound which enters. These compounds are metabolized into a combination of carbon dioxide, oxygen, water, and methane.

Process Description

Waste water is collected from many sources across the plant and sent to a series of digestors. The water is first sent through a screen to collect large solids and is then sent to an anaerobic digestor. In this vessel the soluble solids are consumed along with added nutrients to produce a carbon dioxide and methane mixture, which is sent to the power generation section.

Next the water is pumped to an aerobic digestor where solids are converted to carbon dioxide and water. The flow is then sent to a belt filter press where a small flow of sludge is separated and also sent to the power generation section for combustion. Overall, more than 99% of the

contaminants in the incoming waste water are metabolized and separated from the water, which is now approximately as pure as well water and is then sent throughout the plant as needed.

4.7 Section 7 – Storage

Methodology

The processing plant operates in a steady state manner and therefore there can not be any interruptions due to lack of inputs or lack of volume to store the products. The on-site storage section overcomes this obstacle by providing several days to weeks of inputs and products.

Process Description

The storage section provides on-site storage for the material inputs sulfur dioxide, yeast, enzymes, and fermentation nutrients. It also provides storage for the products of transportation grade ethanol, furfural, turpentine, and a buffer tank to store solids residue before they are combusted.

4.8 Section 8 – Power Generation

Methodology

A significant portion of the wood is not able to be fermented into ethanol and even some of the portions which are able are not fermented due to the less than 100% yields. Furthermore, numerous organic compounds are formed in the pre-treatment and SSF sections that are not beneficial to the process. Therefore, the entire mass of 'solids residue' which exits the solids

recovery stage is combusted on-site to generate processing steam, processing electricity, and excess electricity to be sold to the grid. This section then avoids the need for purchasing steam and electricity, adds another revenue stream by selling green renewable electricity, and avoids the need to dispose of otherwise hazardous organic compounds.

The air emissions from this power generation unit will be comparable to a wood fired boiler plant. The high recovery of sulfur in the pre-treatment section should minimize the release of sulfur oxides, and the only carbon released within this flue gas will have been originally extracted from the atmosphere by the trees, resulting in a closed loop cycle. The full spectrum of air emissions from this combustion are not quantified in this study.

Process description

The solids cake and soluble solids flows from the distillation section, along with the methane and solids sludge flows from waste water treatment, are sent to the power generation section for combustion. These flows enter at a combined lower heating value of approximately 2,900 BTU/lbs and a moisture content of 50%.

This solids residue is combusted in a fluidized bed combustor and produces superheated steam at 1400 psi and 950 F. This steam is passed through an expansion turbine at an isentropic efficiency of 85% and a generator efficiency of 98% to produce electricity and has three steam extraction points. These extraction points at 350, 65, and 25 psi (all pressures quantified in this study are actual pressures, psia, as opposed to gauge pressures) serve to draw off processing

steam for the plant. The steam exits the turbine at a pressure of 1.5 psi where it passes through a cooling tower water condenser under the partial vacuum.

4.9 Section 9 – Utilities

Methodology

The utilities section includes equipment which is necessary for the general operation of the plant.

Process Description

This section includes an air compressor to power the air press in solids recovery as well as an instrument air dryer, the cooling water pump to send water leaving the waste water treatment section to condensers throughout the plant, and a clean in place (CIP) system to periodically clean the fermentors and other equipment.

Chapter 5: Base Case Modeling Results

The base case plant was modeled using the above described sections, components, and assumptions. The resulting plant produced a yield of 79.8 gallons of ethanol per dry ton of pine, or 55,851,200 gallons per year operating 8400 hours per year. The maximum theoretical fermentation yield is 99.9 gallons of ethanol per dry ton of southern pine, assuming only the six carbon sugars are fermented, resulting in a base case yield of 79.9% of theoretical.

The total heat and electrical demand of this base case plant has also been calculated from the model. Since the majority of the energy required by the plant is in the distillation and solids separation area, as discussed previously, this section will receive the majority of the attention when trying to minimize the plant energy demand. The electrical demand of the distillation section has been calculated to be 1,197 kW and the demand of the remaining plant as 11,000 kW.

Processing steam is used in the distillation and solids recovery as well as the pre-treatment sections of the plant. This steam is extracted from the power generation turbine at three different pressures; 350, 65, and 25 psi. The material entering pre-treatment is first heated to 100 C by 65 psi steam, and then raised to 215 C by 350 psi steam. This steam is directly injected into the pre-treatment vessel as live steam as opposed to passing through a heat exchanger. The reboilers of the beer and rectification columns as well as the heat exchanger at the entrance to the molecular sieves are heated by 65 psi steam, and the first effect evaporator is heated by 25 psi steam. The values of these steam flows are shown below in Table 4.

Table 4: Facility Energy Demand for Base Case

	Energy Demand (BTU/hr)	Energy Demand (BTU/gallon EtOH)
Direct Injection Pre-Treatment Steam	112,967,877	16,990.3
Beer Column Reboiler	136,144,990	20,476.2
Rectification Column Reboiler	17,063,412	2,566.3
Evaporator Heat Exchanger	151,924,614	22,849.4
Molecular Sieve Heat Exchanger	1,023,954	154.0
Plant Pump Work	41,625,922	6,260.5
Total	**460,750,768**	**69,296.7**

The energy demand for the directly injected pre-treatment steam is taken as the difference of the enthalpy of the steam from the enthalpy of water at ambient conditions (1 atm, 20 C) multiplied by the flow rate. This results in a relative enthalpy of 1,246 BTU/lbs for the 350 psi steam superheated at 630 F and 1,162 BTU/lbs for the 65 psi steam superheated at 315 F.

The heat demands in the four heat exchangers shown in Table 4 are simply taken as the condensing heat of the steam multiplied by the flowrate. This is conservative, assuming that the steam is only condensing in the heat exchanger giving off its latent heat, but its temperature remains at saturation. A small drop in condensate temperature could be assumed if the heat transfer area were large enough, which in turn would decrease the mass flow rate of steam required.

The listed energy demand for the pump work is simply the electrical load of the equipment, with no correction back to primary energy source. This will be further explored in Results, Section 7.0.

It is shown that the total energy demand for the plant is calculated to be 69,297 BTU per gallon of ethanol. As a reference the lower heating value (LHV) of ethanol is 76,330 BTU per gallon. This would indicate that it takes almost as much energy, not including agricultural inputs, to produce a gallon of ethanol as is in the ethanol itself. However, this entire energy demand is supplied internally through the combustion of the solids residue, and in addition there is enough energy generated from this combustion that excess electricity can be sold to the grid. This indicates that there is no external energy required in the processing of the pine to ethanol, a dramatic improvement over traditional ethanol production facilities.

The amounts of electricity generated, consumed on-site, and sold to the grid in this base case are shown below in Table 5, and the implications of this net energy analysis are discussed in Results, Section 7.0.

Table 5: Base Case Plant Electricity

	Electricity (kW)	**Electricity (kWhr/gallon EtOH)**
Gross Generated	40,968	6.16
Consumed On-Site	12,197	1.83
Excess sold to Grid	28,771	4.33

Chapter 6: Alternate Distillation and Solids Recovery Scenarios

In an attempt to decrease the heat and electrical loads in the distillation and solids recovery section several different configurations and scenarios are considered. These scenarios use a different set of equipment with a different process path, however achieve the same section goals of producing 99.5% w/w pure ethanol and a solids residue at approximately 50% moisture. These scenarios are first rationalized to consider if it is in fact possible for them to reduce the energy demand, if so then they are rigorously modeled and integrated into the entire plant model. Finally, in the Results section (7.0) their change in capital cost is weighed against their change in energy demand to determine the potential payback period.

6.1 Scenario 1 – Pressure Swing Distillation

The first scenario considered is to use a pressure swing distillation setup to skip over the ethanol-water azeotrope instead of using a molecular sieve. This process takes advantage of the fact that the location of the azeotrope is dependent upon the pressure of the mixture. Therefore, the mixture can be purified up to a quality close to the azeotrope, then be pumped to a pressure which forces the azeotrope to a lower purity than what the mixture is currently at, and then the mixture can continue to be purified through conventional distillation. This possibility is explored further in Figure 7 below which shows the ethanol-water equilibrium curves at several elevated pressures.

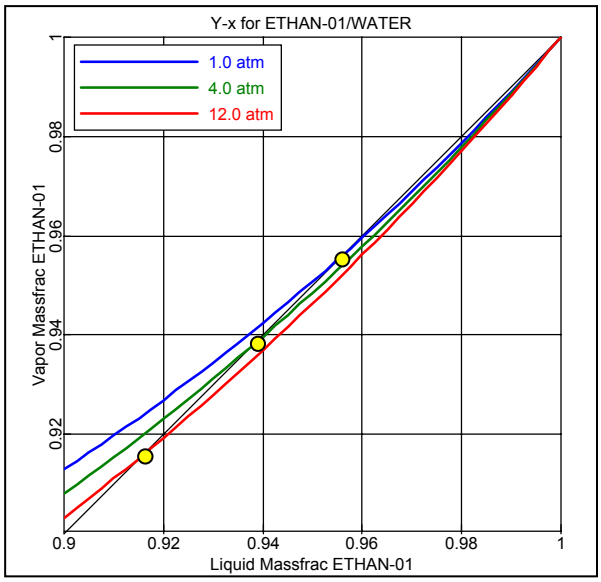

Figure 7: Ethanol-water Equilibrium curves at increased pressures

The yellow dots on the above figure indicate the locations where the equilibrium curves cross the line y=x, and therefore the locations of the azeotrope at the various pressures. This shows that if the ethanol-water mixture is purified up to a quality above 91.5 % w/w through conventional distillation at atmospheric pressure, it can then be pumped to 12 atm to skip over the azeotrope, and then continued to be distilled to a purity of 99.5% w/w without further interference.

This scenario does not appear to be ideal for the separation of ethanol and water due to the location of the azeotrope at such a high quality of ethanol. This does not leave much room for the final purification after the pressure change and the resulting column will only need to be a few stages. The dependence of the location of the azeotrope upon the pressure is also weak for this ethanol-water mixture, meaning that more pump work is needed to move the azeotrope a fixed distance as opposed to other mixtures.

Instead of increasing the pressure of the mixture, the same philosophy can be used and the pressure of the mixture decreased until the azeotrope occurs at a higher quality than the required product. This is to say that as the pressure of the mixture decreases, there will be a point where the ethanol-water azeotrope occurs at ethanol purities over 99.5% w/w and therefore is not a hindrance in this process. This phenomenon is explored further in the below figures examining the ethanol-water equilibrium curves as a function of partial vacuum pressures.

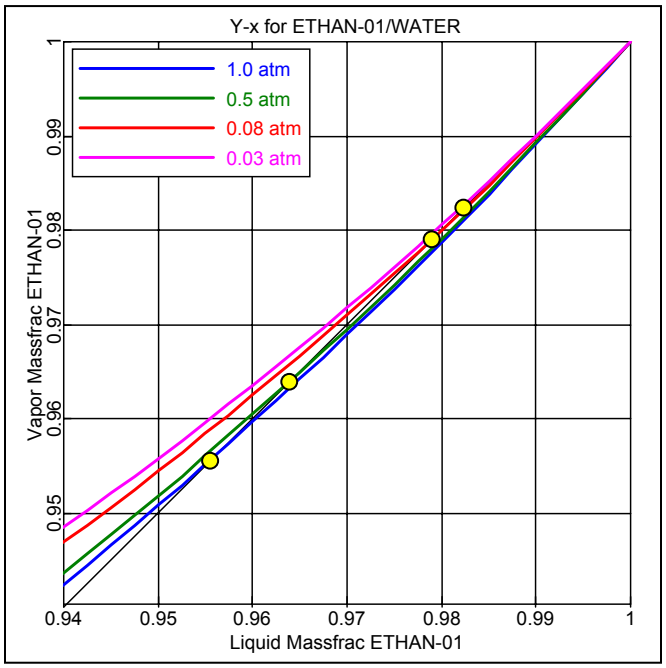

Figure 8: Ethanol-water Equilibrium curves at partial vacuum pressures

This process again does not seem ideal for the separation of ethanol and water. The lowest pressure equilibrium curve shown on the above diagram is at 0.03 atm, which only moves the azeotrope to ~98.3% w/w. An even lower pressure is required to move this azeotrope above 99.5% w/w which would require a high amount of energy, both in the pumps required to force the initial vacuum and in maintaining this vacuum throughout the distillation column.

Although this phenomenon of pressure swing distillation is a powerful tool the specific interactions between the ethanol and water molecules make this separation technique less than ideal, and therefore will not be quantified in this study. The resulting scenario is considered likely to require a higher energy demand as well as a higher capital cost, and therefore is inferior to the base case scenario for this process.

6.2 Scenario 2 – Splitting the Beer Column

A second scenario considered takes advantage of the response of not the mixture azeotrope but the component volatilities to pressure. The base case uses a beer column at atmospheric pressure to purify the ethanol to approximately 30% w/w while also completely splitting off the carbon dioxide and solids flows. This is accomplished by use of a reboiler which operates at a fixed temperature of 100 C and a reflux condenser which takes in approximately 93 C vapor and returns a 60 C liquid.

An alternate configuration could use two beer columns, each of half the capacity as in the base case, however placed in parallel. One of the columns could be pressurized so that the temperature of the fluid exiting the condenser is greater than the temperature of the reboiler in the other column. This would allow for the condensing energy of the pressurized column to be used as reboiler energy in the atmospheric column.

The pressure of the second column is determined by the need for the flow leaving the condenser of this second column to be a liquid. This condenser is in contact with the reboiler of the

atmospheric column, which operates at 100 C. Therefore, there needs to be a large enough temperature difference between the liquid flow leaving the condenser of the pressurized column and the 100 C reboiler of the atmospheric column. This minimum temperature difference is set to 20 degrees C, and therefore the pressure is found which has a saturation temperature of the more volatile component, ethanol, at 120 C.

This pressure is 4.2 atm, and the equilibrium curves for an ethanol-water mixture at this pressure, as well as at atmospheric, are shown below in Figure 9. Note that the saturation temperature for pure ethanol (right axis) is 120 C, and it also shows that since we expect the concentration of the ethanol leaving this column to be approximately 30% w/w that this vapor flow will be at approximately 140 C.

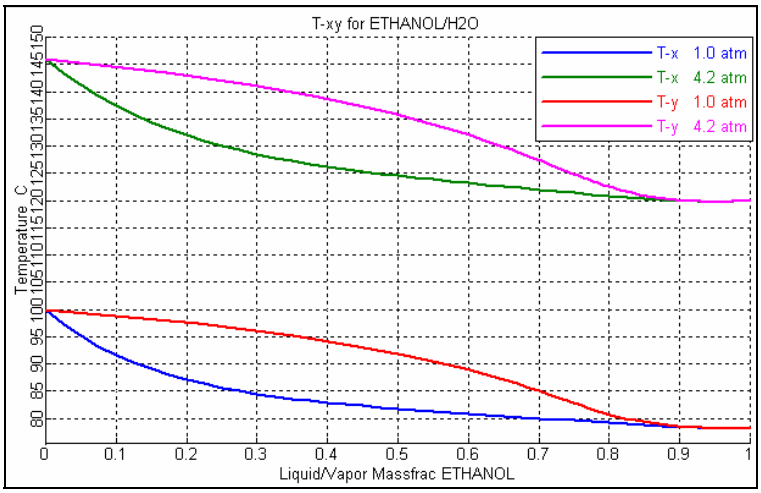

Figure 9: Ethanol-Water Equilibrium Curves at 1 and 4.2 atm

McCabe-Thiele diagrams were constructed to determine the minimum required number of stages and reflux ratio of this new pressurized column, however the presence of the solids flow as well as the remaining gaseous carbon dioxide within the mixture made this process non-ideal. As a

result, the column was kept to 16 stages and a molar reflux ratio of 3 as in the atmospheric column in the base case. The mass flow rates of the ethanol vapor side draw and the bottoms were varied to produce a bottoms flow with an ethanol mass fraction of 0.0005 and a condenser temperature below 120 C.

A schematic of this configuration as well as the numerical results of the heat loads are shown graphically below in Figure 10.

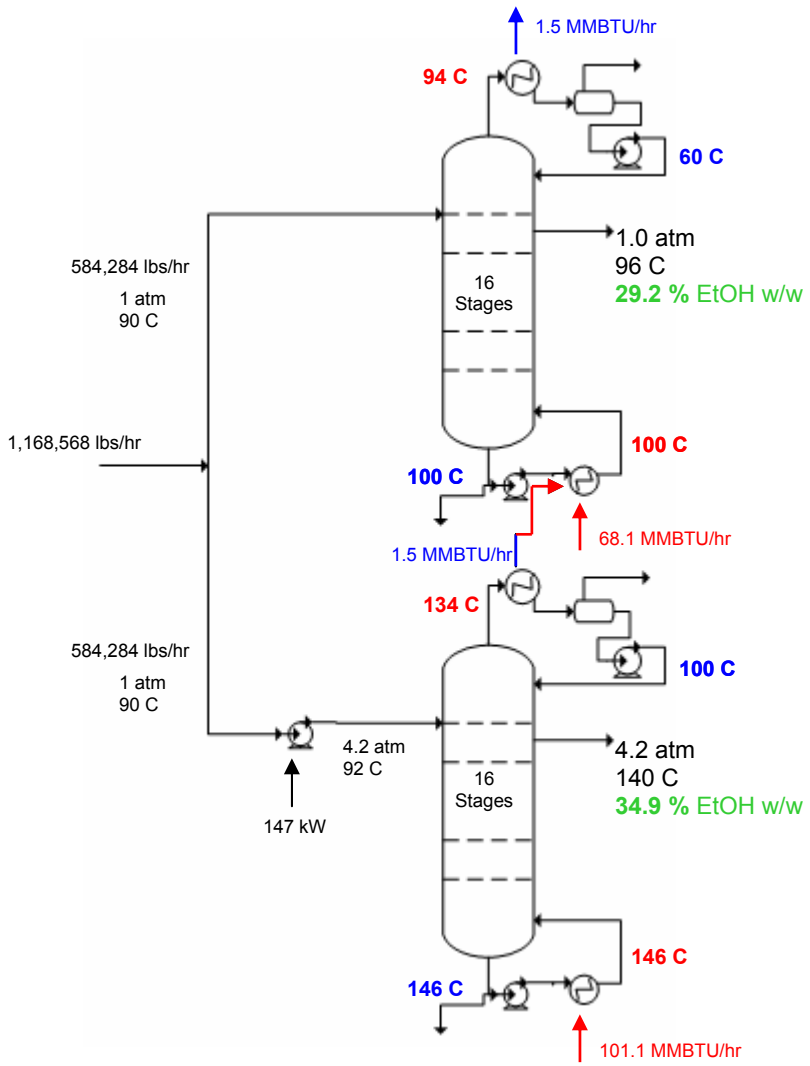

Figure 10: Schematic of Scenario 2
(red values indicate a vapor, blue values liquid)

It can be seen in the above figure that the condenser of the pressurized vessel takes a 134 C vapor and condenses it to a 100 C liquid, while the reboiler of the atmospheric column operates at a constant 100 C. This presents an opportunity for heat integration, and it can be seen that the 1.5 MM BTU/hr condensing load of the pressurized column is used as a heat source for the reboiler of the atmospheric column.

At first glance it appears that this scenario in fact has a higher heat load than the base case, since the combined heat loads of the reboilers is 167.7 MM BTU/hr, which includes the -1.5 MM BTU/hr heat integration subtraction, as opposed to 136.1 MM BTU/hr load in the base case. However, it must also be noticed that the ethanol vapor and bottoms flows leave at the elevated pressure of 4.2 atm and also the elevated temperatures of 140 and 146 C, respectively. Furthermore, the ethanol vapor leaves the pressurized column at a concentration of 34.9% as opposed to 29.2% w/w in the base case. To fully understand the implications of this configuration it must be integrated with the rest of the distillation and solids recovery section to quantify the total heat and electrical load. This was performed and the resulting energy loads are presented below in Table 6.

Table 6: Facility Energy Demand for Scenario 2

	Energy Demand (BTU/hr)	**Energy Demand (BTU/gallon EtOH)**
Direct Injection Pre-Treatment Steam	112,967,877	16,990.3
Net Beer Column Reboilers	167,695,373	25,221.3
Rectification Column Reboiler	14,084,149	2,118.3
Evaporator Heat Exchanger	75,521,019	11,358.3
Molecular Sieve Heat Exchanger	1,026,789	154.4
Plant Pump Work	42,219,749	6,349.8
Total	**413,514,956**	**62,192.5**

The above table shows that while the net heat loads in the two beer columns is greater than the heat load in the single beer column in the base case, the total energy demand is lower in this split column scenario. The decreased heat load in the rectification column as well as the dramatic decrease in the heat load of the evaporator heat exchanger are greater than the increase in energy demand in the corresponding beer columns. As expected, the pre-treatment direct injection steam and the molecular sieve heat exchanger loads remain unaffected.

While the heat load is decreased in this scenario, it is also noticed that the electrical load has slightly increased. The amount of electricity generated, consumed on-site, and sold to the grid in this configuration is summarized below in Table 7. It is also noted that this setup requires additional equipment as compared to the base case. This increased capital, as well as the increased electrical load and decreased heat load, will be compared and discussed in Results, Section 7.0.

Table 7: Scenario 2 Plant Electricity

	Electricity (kW)	**Electricity (kWhr/gallon EtOH)**
Gross Generated	42,627	6.41
Consumed On-Site	12,371	1.86
Excess sold to Grid	30,256	4.55

6.3 Scenario 3 – Mechanical Vapor Recompression

A third scenario considered to reduce the total energy demand of this facility is to use mechanical vapor recompression (MVR) to re-use low quality waste steam as a heat input. Low quality steam, which is currently being simply condensed with cooling water and used as recycle water, can be recompressed with a mechanical vapor compressor and raised to a usable quality.

The target low quality waste steam in this facility is the low pressure water vapor leaving the third effect evaporator. Currently this flow is being condensed with cooling water and then sent to the first SSF vessel to act as dilution water to achieve the desired solids loading. However, with the input of mechanical energy this steam can be compressed to a high enough quality to be used as a heat input to the first effect evaporator.

The first effect evaporator operates at 83.4 C and is heated by primary steam at 25 psi and 112 C as well as the condensing load from the rectification column at 93 C. The water vapor flow leaving the third effect evaporator is saturated at 0.211 atm and 63.0 C. For this flow to be able to be used as a heat source for the first effect evaporator it must have a temperature some delta T above 83.4 C. This flow must also be at a high enough pressure so that the vapor will condense at this temperature, thus transferring its latent heat to the evaporator. The compression of this flow in a steam re-compressor handles both of these requirements, significantly increasing the temperature while also increasing the pressure.

The minimum required temperature difference between the two flows is taken to be 10 C. This means that the water vapor must be compressed to a pressure at or above the saturation pressure of water at 93.4 C, which is calculated to be 0.80 atm (11.8 psi). Electrical energy will be required to compress this flow from 0.211 to 0.80 atm, however the full condensing heat of this flow will then be available for heat integration. This is illustrated graphically in the temperature-entropy diagram of water as Figure 11 shown below.

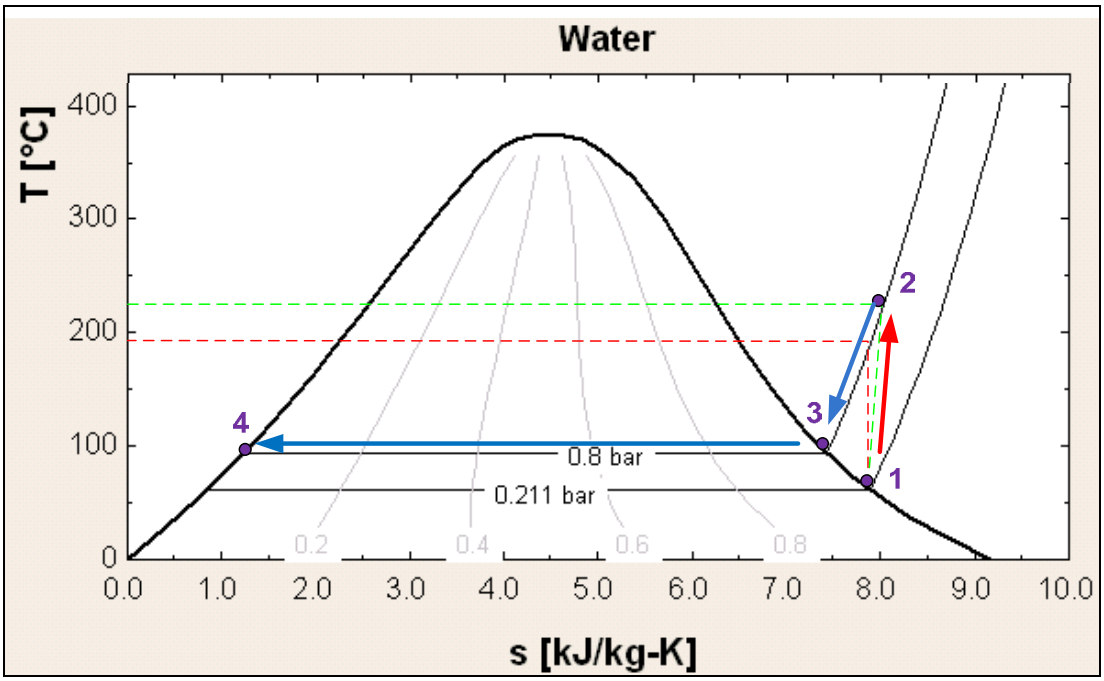

Figure 11: T-s diagram of Water

The trade-off is that electrical energy is put in to the steam as it is compressed from points 1 to 2 (red arrow), while heat is given off as this steam cools to a saturated vapor and then to a saturated liquid, points 2 to 3 to 4 (blue arrows). For this process to be beneficial the electrical energy input to the steam, multiplied by a scaling factor to take it back to primary heat energy, must be less than the amount of heat energy released when the steam condenses.

The red line on the above image shows the isentropic compression of water at 0.211 atm as a saturated vapor to a superheated vapor at 0.80 atm, with the temperature increasing from 63 C to 191 C. The specific work calculated for this compression is 242.9 kj/kg. The green line shows the same compression, however at an isentropic efficiency of 80%, resulting in a superheated vapor at 221 C, and a specific work of 303.6 kj/kg.

To analyze the ability of MVR to decrease the energy demand of this facility a portion of the water vapor flow leaving the third effect evaporator was compressed to 0.80 atm at an isentropic efficiency of 80%. Only enough vapor was compressed so as to fully supply the heat load of the first effect evaporator, with an assumed heat loss of 4%. This required that 54.5% of the exiting water vapor be compressed and required 4,963 kW of electrical energy.

The condensing heat of this compressed vapor, as well as the condensing heat of the condenser of the rectification column, completely supply the heat load of the evaporator and avoid the need for primary steam to be diverted from the power generation turbine. The new total energy demand for this facility is shown below in Table 8.

Table 8: Facility Energy Demand for Scenario 3

	Energy Demand (BTU/hr)	**Energy Demand (BTU/gallon EtOH)**
Direct Injection Pre-Treatment Steam	112,967,877	16,990.3
Beer Column Reboiler	136,144,990	20,476.2
Rectification Column Reboiler	17,063,412	2,566.3
Evaporator Heat Exchanger	0	0
Molecular Sieve Heat Exchanger	1,023,954	154.0
Plant Pump Work	58,563,648	8,808.0
Total	**325,763,881**	**48,994.8**

The above table shows that the use of MVR has reduced the energy demand of the facility from 460.8 to 325.8 MM BTU/hr, a 29% reduction. However, it must also be noted that electrical energy has a greater value than primary heat energy because of the inefficiencies inherent in being produced, and therefore this energy comparison must be further evaluated.

As far as an energy analysis, the amount of additional electrical energy input (4,963 kW = 16,938,700 BTU/hr) is far outweighed by the amount of condensing heat received as an output (151,924,614 BTU/hr). The trade-off of this steam versus electrical energy can best be quantified by examining the amount of electricity generated, consumed, and sold as excess in this plant. Through the rigorous calculation of the electrical generation a factor can be back calculated for each quality of steam which relates the value of the primary heat energy to the electrical energy. The electricity summary for this scenario is shown below in Table 9.

Table 9: Scenario 3 Plant Electricity

	Electricity (kW)	Electricity (kWhr/gallon EtOH)
Gross Generated	48,073	7.23
Consumed On-Site	17,160	2.58
Excess sold to Grid	30,913	4.65

The above table shows that more electricity is available to be sold to the grid in this scenario than in the base case (30,913 vs. 28,771 kW). This means that more electricity has been generated due to the decreased steam demand than has been consumed by the increased electrical load due to the steam compressor. This summary shows that in fact this scenario has decreased the total plant energy demand with respect to the base case.

However, this scenario also requires additional equipment that was not included in the base case, which means that this will have a higher capital cost. The additional capital cost, as well as its implications towards the added value of this scenario, will be evaluated in a cost-benefit analysis in Results, Section 7.0.

6.4 Scenario 4 – Mechanical Vapor Recompression, all Vapor

The recompression of the waste low quality steam to 0.80 atm to provide 93 C steam to the first effect evaporator has been shown to be beneficial. The next step is to see if the remainder of the steam should be compressed to the next highest quality heat load to further decrease the primary steam demand. The next lowest quality heat load is in the reboiler of the beer column, which operates at 100 C.

Using the same reasoning as in Scenario 3, the 0.211 atm steam exiting the third effect evaporator must be compressed to 1.4 atm. This will provide a 10 C temperature difference between a saturated liquid at 1.4 atm and the 100 C heat load. Again assuming an 80% isentropic compressor efficiency, it will require 465.8 kj/kg of electrical energy to compress the steam, which will be superheated to 306 C during the process.

The remaining 45.5% of the vapor flow, after being compressed to 1.4 atm, provides 122.4 MM BTU/hr of heat energy at 110 C when condensed in the beer column reboiler, assuming a 4% heat loss. The compression of this flow requires 6,316 kW of electrical energy. As in Scenario 3, the benefit of this trade-off is best quantified by comparing the amount of excess electricity available for sale. This takes into account the decreased steam load as well as the increased electrical load, with the effective multiplier attached to convert the electrical energy back to primary heat energy. This information is presented below in Tables 10 and 11.

Table 10: Facility Energy Demand for Scenario 4

	Energy Demand (BTU/hr)	Energy Demand (BTU/gallon EtOH)
Direct Injection Pre-Treatment Steam	112,967,877	16,990.3
Beer Column Reboiler	18,649,876	2,804.9
Rectification Column Reboiler	17,063,412	2,566.3
Evaporator Heat Exchanger	0	0
Molecular Sieve Heat Exchanger	1,023,954	154.0
Plant Pump Work	80,118,893	12,049.9
Total	**229,824,011**	**34,565.4**

Table 11: Scenario 4 Plant Electricity

	Electricity (kW)	Electricity (kWhr/gallon EtOH)
Gross Generated	55,946	8.41
Consumed On-Site	23,476	3.53
Excess sold to Grid	32,470	4.88

While it is noted that the total plant energy demand has been decreased in this scenario with respect to Scenario 3, as well as the base case, it is more important to note that the amount of excess electricity sold to the grid has increased with respect to these two cases. This shows that the decreased heat load on the power generation turbine has outweighed the increased electrical load on this turbine, even after the electrical load is corrected back to its primary heat energy.

6.5 Scenario 5 – Split Beer Column and very low pressure MVR

The energy saving benefits discussed in Scenarios 2 and 3 are not mutually exclusive, since Scenario 2 deals solely with the beer column and Scenario 3 solely with the evaporator. Therefore, under this scenario both modifications are made, combining the benefits of the split beer column with the mechanical vapor recompression of the water vapor leaving the evaporator.

Under this scenario the splitting of the beer columns works exactly as in Scenario 2, however the dynamics within the three effects of the evaporator are different. Since a higher temperature and pressure flow is being sent to the first effect evaporator than in the base case, the amount of water vapor leaving each effect has changed. In the resulting configuration, there is a dramatically increased water vapor flow leaving the third effect evaporator which is available for MVR (238,000 vs. 73,500 lbs/hr), while still at the same temperature and pressure as in the base case.

This results in the complete supply of heat to the first effect evaporator with only 22.3% of the exiting vapor flow being compressed. This presents an opportunity to compress the remaining vapor flow to fulfill the heat requirements of other components in the plant. The remaining heat exchanger loads for components in the facility are shown below in Table 12. Along with the heat loads are the operating temperatures and the required steam pressure to supply heat to this load, which is the saturation pressure of water at the specified operating temperature plus 10 C.

Table 12: Remaining Plant Heat Exchanger Loads

Component	**Heat Demand** (MMBTU/hr)	**Operating Temperature** (C)	**Required Steam Pressure** (atm)
Pressurized Beer Column	101.1	146	5.6
Rectification Column	14.1	121	2.8
Molecular Sieve	1.0	116	2.4
Atmospheric Beer Column	66.6	100	1.4

It is shown that steam is needed at a variety of pressures ranging from 1.4 to 5.6 atm to supply all of the remaining heat loads. Due to the low relative heat demands and relatively close required steam pressure, both the rectification column and the molecular sieve will be supplied by 2.8 atm steam. In a facility the recompression to these three pressures could represent three separate

compressors, or a single compressor with steam extraction ports at these three specified pressures.

The compressor outlet temperature and specific work increase with increased outlet steam pressure, however the condensing heat being transferred to the medium remains relatively constant. This relation is explored graphically below in Figure 12. This figure shows the recompression of steam with an incoming pressure of 0.211 atm and temperature of 63 C being compressed at an 80% isentropic efficiency to the pressures shown on the x-axis. The right axis shows the condensing heat returned from the recompressed steam when sent through the heat exchanger, and the left axis shows the electrical work required to compress the steam to the specified pressure. Although it is shown that the electrical energy is far less than the condensing heat energy, it must be noted that this electrical energy is not corrected back to a primary or heat energy value. This correction takes place in the calculation of the total excess electricity available for sale in each scenario.

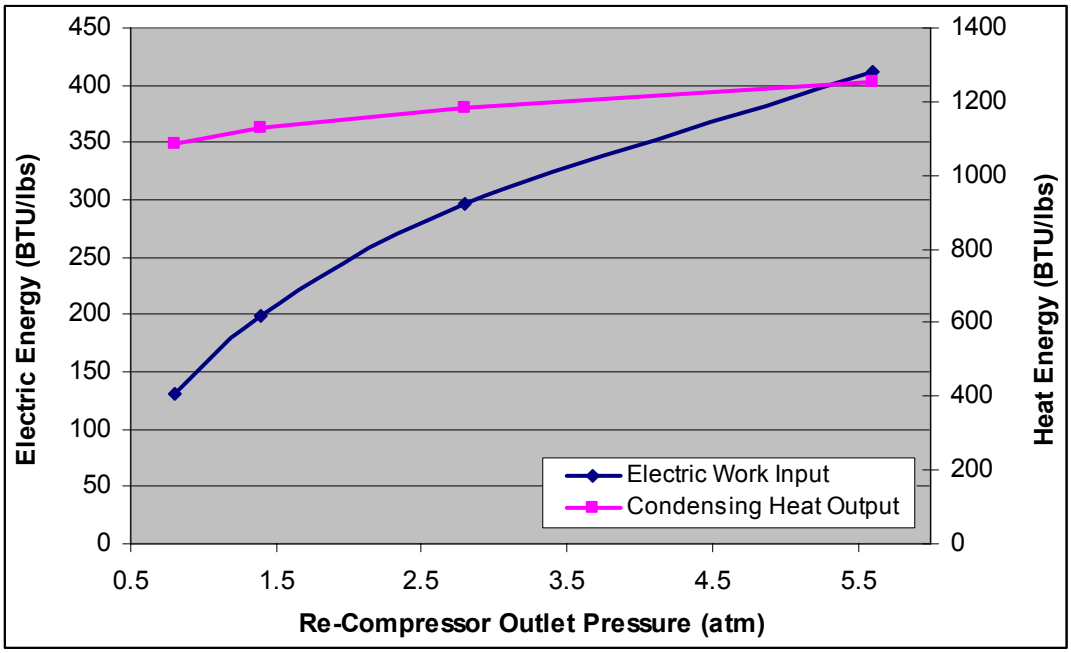

Figure 12: Relation of Input Electrical Energy to Returned Condensing Heat Energy

The compression of the 0.211 atm steam to 0.8 atm requires 303.6 kj/kg (130.5 BTU/lbs), to 1.4 atm requires 465.8 kj/kg (200.3 BTU/lbs), to 2.8 atm requires 691.5 kj/kg (297.3 BTU/lbs), while the 5.6 atm steam uses 958.9 kj/kg (412.3 BTU/lbs), all at 80% isentropic efficiency. As a result of this strong dependence this configuration will be analyzed in stages. In this scenario only enough vapor will be recompressed to supply the heat load of the first effect evaporator, and the rest will simply be condensed with cooling water as in the base case. In the next scenario the evaporator as well as the atmospheric beer column will be heated by re-compressed steam, the loads supplied by the lower pressure 2.8 atm steam will be analyzed in the following scenario, and then the incorporation of the 5.6 atm steam will be included in the following scenario.

The temperature-entropy diagrams for water vapor under the two higher compressions are shown below in Figure 13 to better demonstrate the strong dependence of the specific work and compressor outlet temperature on the required steam pressure.

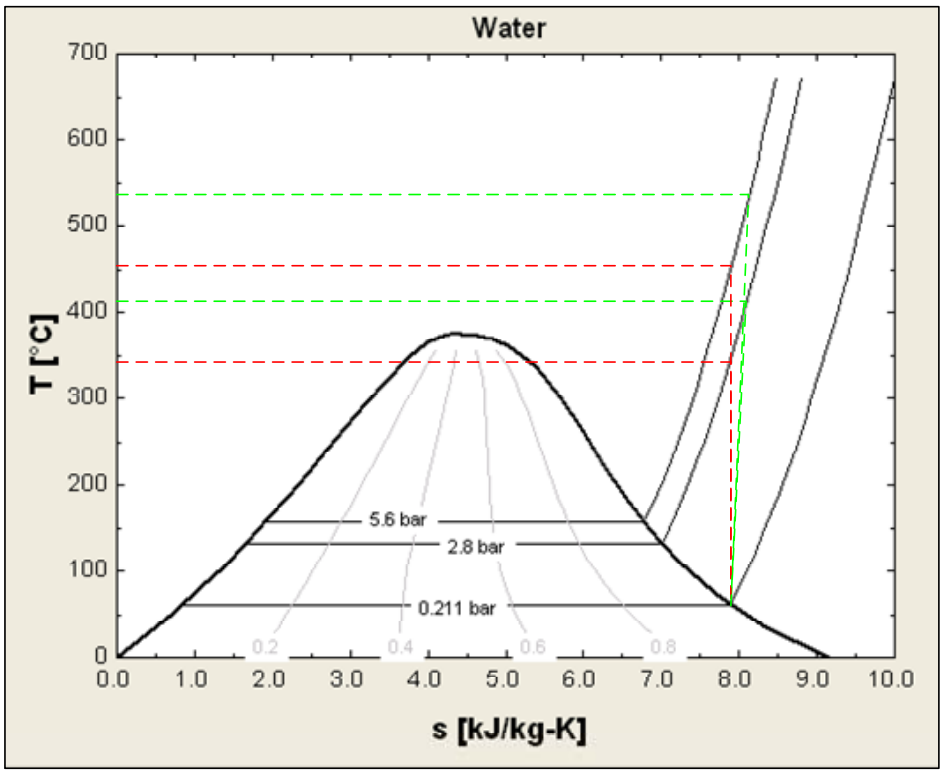

Figure 13: T-s diagram of water

As stated previously, 22.3% of the low pressure water vapor is recompressed to fully supply the heat to the first effect evaporator, assuming a 4 % heat loss in the heat exchanger. The electrical load on this compressor is 2,023 kW. The total plant energy demand for this scenario is shown below in Table 13.

Table 13: Facility Energy Demand for Scenario 5

	Energy Demand (BTU/hr)	Energy Demand (BTU/gallon EtOH)
Direct Injection Pre-Treatment Steam	112,967,877	16,990.3
Beer Column Reboiler	167,695,373	25,221.3
Rectification Column Reboiler	14,084,149	2,118.3
Evaporator Heat Exchanger	0	0
Molecular Sieve Heat Exchanger	1,026,789	154.4
Plant Pump Work	49,123,843	7,388.2
Total	**344,898,031**	**51,872.5**

The steam demand for this combined configuration is in fact decreased from the split beer column configuration in Scenario 2, however it is slightly increased from the mechanical vapor recompression configuration in Scenario 3. The electrical load in this scenario is significantly lower than in Scenario 3, yet the electrical load in this scenario is increased when compared to Scenario 2. Once again the trade-off of decreased heat versus increased electrical demand must be analyzed through the amount of excess electricity available for sale. The electrical generation, consumption, and excess sale flows for this scenario are shown below in Table 14.

Table 14: Scenario 5 plant Electricity

	Electricity (kW)	Electricity (kWhr/gallon EtOH)
Gross Generated	46,158	6.94
Consumed On-Site	14,394	2.16
Excess sold to Grid	31,764	4.78

The amount of excess electricity able to be sold to the grid is higher in this scenario than in either of the individual configurations, however lower than in Scenario 4 where re-compressed steam also supplied heat to the beer column. Furthermore, this scenario requires more equipment than Scenarios 2 and 3 and a comparable amount to Scenario 4. This relationship of trade-offs will be discussed further in Results, Section 7.0.

6.6 Scenario 6 – Split Beer Column and very low and low pressure MVR

This scenario involves the same process as in Scenario 5, however re-compresses steam to supply heat to the atmospheric beer column (at 1.4 atm) as well as the first effect evaporator (at

0.80 atm). The total heat load of the atmospheric beer column is 66.6 MM BTU/hr and requires 25.9% of the water vapor leaving the third effect evaporator to be compressed to 1.4 atm, assuming a 4% heat loss. This compression requires 3,580 kW of electrical energy.

The plant energy demand and electricity profile under this scenario are shown below in Tables 15 and 16

Table 15: Facility Energy Demand for Scenario 6

	Energy Demand (BTU/hr)	Energy Demand (BTU/gallon EtOH)
Direct Injection Pre-Treatment Steam	112,967,877	16,990.3
Beer Column Reboiler	101,095,373	15,204.7
Rectification Column Reboiler	14,084,149	2,118.3
Evaporator Heat Exchanger	0	0
Molecular Sieve Heat Exchanger	1,026,789	154.4
Plant Pump Work	61,341,667	9,225.8
Total	**290,515,855**	**43,693.5**

Table 16: Scenario 6 plant Electricity

	Electricity (kW)	Electricity (kWhr/gallon EtOH)
Gross Generated	50,621	7.61
Consumed On-Site	17,974	2.70
Excess sold to Grid	32,647	4.91

This scenario has the lowest total plant energy demand of all the scenarios previously analyzed, with the exception of Scenario 4, however it more importantly has the highest amount of excess electricity available for sale to the grid. This shows that the additional electricity required to compress the steam to 1.4 atm was outweighed by the amount of primary energy displaced by this heat integration.

6.7 Scenario 7 – Split Beer Column and very low, low, and medium pressure MVR

This setup is the same as in Scenario 6, however it also includes recompression of steam to 2.8 atm to supply heat to the rectification column and molecular sieve. An additional 5.6% of the water vapor leaving the third effect evaporator is needed to fully supply this combined heat load of 15.1 MM BTU/hr at 121 C, assuming a 4% heat loss. The compression of this flow requires 1,153 kW of electrical energy.

The effects that this additional vapor compression has on the total plant energy demand as well as the amount of electricity generated, consumed on-site, and excess sold to the grid are shown in Tables 17 and 18 below.

Table 17: Facility Energy Demand for Scenario 7

	Energy Demand (BTU/hr)	Energy Demand (BTU/gallon EtOH)
Direct Injection Pre-Treatment Steam	112,967,877	16,990.3
Beer Column Reboiler	101,095,373	15,204.7
Rectification Column Reboiler	0	0
Evaporator Heat Exchanger	0	0
Molecular Sieve Heat Exchanger	0	0
Plant Pump Work	65,276,626	9,817.6
Total	**279,339,876**	**42,012.6**

Table 18: Scenario 7 plant Electricity

	Electricity (kW)	Electricity (kWhr/gallon EtOH)
Gross Generated	51,633	7.77
Consumed On-Site	19,127	2.88
Excess sold to Grid	32,506	4.89

The above tables show that while this scenario has a decreased plant energy demand and increased amount of electricity generated with respect to Scenario 6, it has less excess electricity

available to be sold to the grid. This shows that the decreased steam load was not enough to outweigh the additional electrical load of the compression of the steam to 2.8 atm. Furthermore, this scenario has additional equipment than Scenario 6, and therefore a higher capital cost. With less electricity available for sale and an increased capital cost, it is not possible for this scenario to be beneficial with respect to Scenario 6.

6.8 Scenario 8 – Split Beer Column and total MVR

This scenario considers the compression of the water vapor leaving the third effect evaporator to 5.6 atm to be used to supply the heat load of the pressurized beer column. This compression requires a tremendous amount of energy, 958.9 kj/kg (412.3 BTU/lbs), however it has the potential to offset this consumption through additional electricity generation by decreased medium pressure steam demand.

In this scenario 46.2% of the vapor flow is already being re-compressed to supply other heat loads, yet it is calculated that to fully supply the heat load of the pressurized column it doesn't require the entire remainder of the flow. Instead, only 35.4% of the flow is needed, and since this is the last of the heat loads in the entire facility the remaining 10.8% must simply be condensed by cooling water and used as recycle water throughout the plant. The only remaining thermal load in the plant is the direct injection steam in the pre-treatment vessels. This steam is at 23 atm (350 psi) and must be taken directly from the power generation turbine as opposed to the recompression of lower quality steam.

The total heat load of the pressurized beer column of 101.1 MM BTU/hr is supplied by this recompressed steam, assuming a 4% heat loss. This compression of the steam to 5.6 atm requires 10,121 kW of electrical energy. The total plant energy and electricity portfolios under this configuration are shown below in Tables 19 and 20.

Table 19: Facility Energy Demand for Scenario 8

	Energy Demand (BTU/hr)	Energy Demand (BTU/gallon EtOH)
Direct Injection Pre-Treatment Steam	112,967,877	16,990.3
Beer Column Reboiler	0	0
Rectification Column Reboiler	0	0
Evaporator Heat Exchanger	0	0
Molecular Sieve Heat Exchanger	0	0
Plant Pump Work	99,817,574	15,012.5
Total	**212,785,452**	**32,002.9**

Table 20: Scenario 8 plant Electricity

	Electricity (kW)	Electricity (kWhr/gallon EtOH)
Gross Generated	58,407	8.78
Consumed On-Site	29,248	4.40
Excess sold to Grid	29,159	4.39

In a continuing trend, this scenario has the lowest total plant energy demand, generates the most electricity, but has the least amount of excess electricity available to be sold to the grid of all the previously analyzed scenarios, except for the base case. The reasoning is also the same; while this configuration eliminates all steam extraction from the power generation turbine except for the pre-treatment steam, it also has the greatest on-site electrical consumption which outweighs this decreased steam demand. This trade-off, as well as the trade-off with capital cost, is discussed further in Results, Section 7.0.

Chapter 7: Results & Discussion

In this study seven alternate scenarios have been developed and quantified to attempt to reduce the energy demand of an ethanol production facility. Each scenario is evaluated with respect to a base case facility. The energy demands have been quantified in terms of direct injection steam at 350 and 65 psi for pre-treatment, heat exchanger steam at 65 psi for the beer column, rectification column, and molecular sieve, 25 psi steam for the evaporator, and plant electrical load. These values for the base case and all seven alternate scenarios are shown in Tables 21 and 22, and presented graphically in Figure 14 below.

Table 21: Summary of Facility Energy Demands for all Scenarios (MM BTU/hr)

		Alternate Scenario						
	Base Case	**2**	**3**	**4**	**5**	**6**	**7**	**8**
Direct Injection PT Steam	113.0	113.0	113.0	113.0	113.0	113.0	113.0	113.0
Beer Column Reboiler	136.1	167.7	136.1	18.6	167.7	101.1	101.1	0.0
Rect. Column Reboiler	17.1	14.1	17.1	17.1	14.1	14.1	0.0	0.0
Evaporator HX	151.9	75.5	0.0	0.0	0.0	0.0	0.0	0.0
MS HX	1.0	1.0	1.0	1.0	1.0	1.0	0.0	0.0
Pump Work	41.6	42.2	58.6	80.1	49.1	61.3	65.3	99.8
Total	**460.8**	**413.5**	**325.8**	**229.8**	**344.9**	**290.5**	**279.3**	**212.8**

Table 22: Summary of Facility Energy Demands for all Scenarios (BTU/gallon EtOH)

		Alternate Scenario						
	Base Case	**2**	**3**	**4**	**5**	**6**	**7**	**8**
Direct Injection PT Steam	16,990	16,990	16,990	16,990	16,990	16,990	16,990	16,990
Beer Column Reboiler	20,476	25,221	20,476	2,805	25,221	15,205	15,205	0
Rect. Column Reboiler	2,566	2,118	2,566	2,566	2,118	2,118	0	0
Evaporator HX	22,849	11,358	0	0	0	0	0	0
MS HX	154	154	154	154	154	154	0	0
Pump Work	6,261	6,350	8,808	12,050	7,388	9,226	9,818	15,013
Total	**69,297**	**62,192**	**48,995**	**34,565**	**51,873**	**43,693**	**42,013**	**32,003**

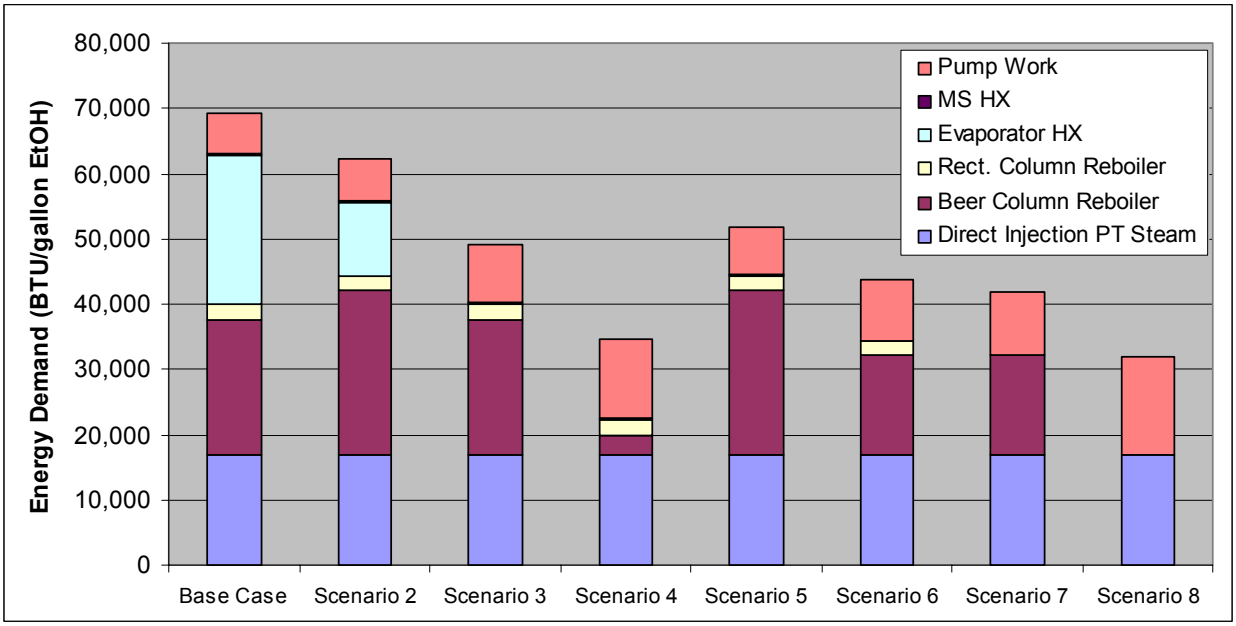

Figure 14: Bar graph of energy demand for all scenarios

From the information presented in this energy demand summary it would appear that Scenarios 4 and 8 require the least amount of energy to produce a gallon of ethanol. However, as mentioned previously, the pump work shown is not corrected back to its primary heat value. Therefore, a scalar must be added to the pump energy which can distort the results when viewed from simply an end use energy standpoint.

Secondly, the energy demand shown here only includes the condensing heat of the steam which is transferred through the heat exchanger. This does not take into account the temperature of this heat exchange, which requires higher quality steam to transfer heat to a higher temperature medium. This also can distort the results from this viewpoint, since steam is extracted at 65 and 25 psi to be used in heat exchangers, as well as 350 psi for pre-treatment. This shows that heat can be transferred to the evaporator from a lower quality steam than what is required to heat the beer column, rectification column, and molecular sieve.

A rigorous way to correct for these two inaccuracies is to examine not the total plant energy demand, but the amount of excess electricity available to be sold to the grid after the on-site consumption has already been subtracted. The power generation turbine already rigorously takes into account the mass flow and quality of all steam extracted when calculating the power generation. It also uses the built in inefficiencies, 85% isentropic turbine efficiency and 98% generator efficiency, to correct the electricity use back to primary heat energy.

As a result, the scenario which has the most excess electricity available for sale to the grid, after subtracting all on-site utilization, is truly the configuration with the lowest plant energy demand. This information is presented below in Tables 23 and 24, and graphically in Figure 15.

Table 23: Plant Electricity for all Scenarios (kW)

		Alternate Scenarios						
	Base Case	**2**	**3**	**4**	**5**	**6**	**7**	**8**
Gross Electricity Generated	40,968	42,627	48,073	55,946	46,158	50,621	51,633	58,407
On-Site Consumption	12,197	12,371	17,160	23,476	14,394	17,974	19,127	29,248
Net Electricity Sold	**28,771**	**30,256**	**30,913**	**32,470**	**31,764**	**32,647**	**32,506**	**29,159**

Table 24: Plant Electricity for all Scenarios (kWhr/gallon EtOH)

		Alternate Scenarios						
	Base Case	**2**	**3**	**4**	**5**	**6**	**7**	**8**
Gross Electricity Generated	6.16	6.41	7.23	8.41	6.94	7.61	7.77	8.78
On-Site Consumption	1.83	1.86	2.58	3.53	2.16	2.70	2.88	4.40
Net Electricity Sold	**4.33**	**4.55**	**4.65**	**4.88**	**4.78**	**4.91**	**4.89**	**4.39**

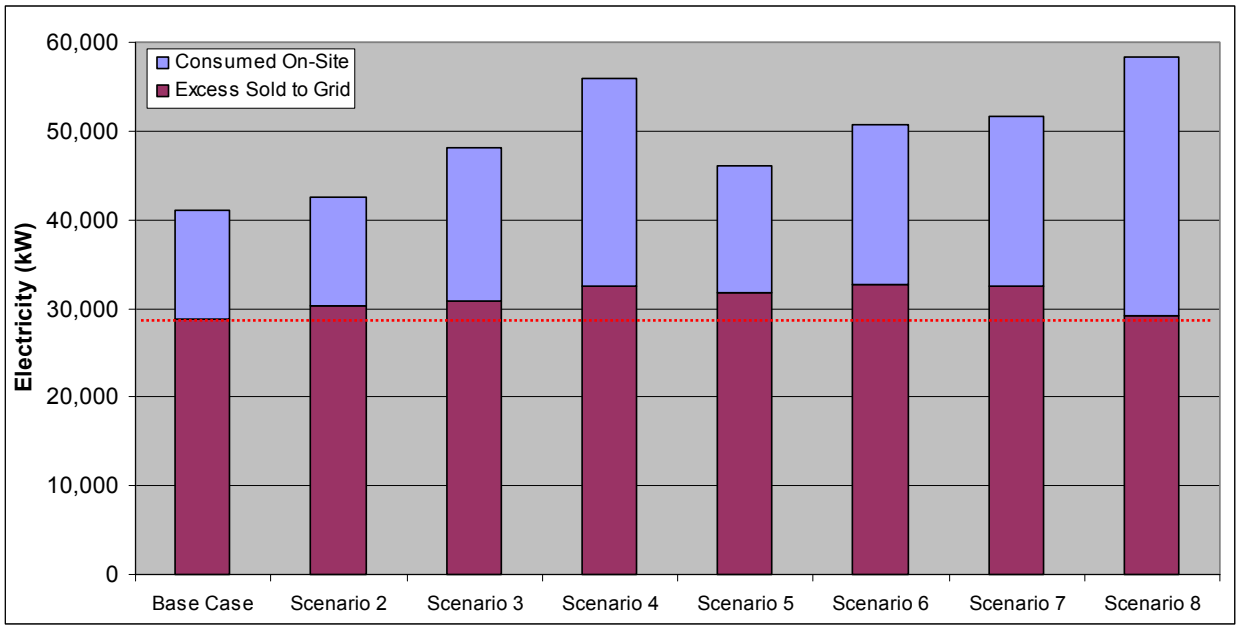

Figure 15: Bar graph of plant electricity for all Scenarios (kW)

The above figure shows that every scenario evaluated has more excess electricity which can be sold to the grid than the base case. There is an obvious trend of increasing electricity generation as the steam demand decreases. This occurs from the base case to Scenario 3, and then to Scenario 4 as more waste steam is sent through MVR. This also occurs in the progression from Scenario 2 to 5, to 6, to 7, to 8. However, while total electricity generation is increasing the amount of electricity consumed on-site is also increasing.

This trade-off of increasing generation vs. increasing consumption reaches an optimum between the re-compression of waste steam to 1.4 and 2.8 atm. This is to say that whenever waste steam is compressed to 1.4 atm (Scenarios 4 and 6), the decreased primary steam consumption leads to an increased electricity generation which outweighs the additional on-site electricity consumption. However, whenever waste steam is compressed to 2.8 atm (Scenario 7), the

decreased primary steam consumption leads to an increased electricity generation which *does not* outweigh the additional on-site electricity consumption.

This trend results in the most excess electricity available for sale to the grid to occur in Scenario 6, where steam is recompressed to 1.4 atm but not to 2.8 atm.

The splitting of the beer column is shown to be beneficial on all counts. The split beer column configuration produces more excess electricity than the base case (Scenario 2 vs. base case), the splitting of the beer column and recompression of waste steam to 0.8 atm produces more excess electricity than recompressing the steam alone (Scenario 5 vs. Scenario 3), and the splitting of the beer column and the recompression of waste steam to 0.8 and 1.4 atm produces more excess electricity than recompressing the steam alone (Scenario 4 vs. Scenario 6).

As mentioned in the modeling section, the increase in revenue from the increase in excess electricity must be weighed against any increases in capital cost. The Aspen Tech software Aspen Icarus 2004.1[3] was used to determine the increased capital cost of a single beer column versus two beer columns, each of half capacity, with one pressurized to 4.2 atm. This software directly uses the output from the Aspen Plus 2004.1 model created in the modeling section, and as a result the equipment costed is directly sized for this facility.

Aspen Icarus 2004.1 calculated an installed cost of $1,183,200 for the beer column in the base case, $796,400 for the atmospheric and $980,700 for the pressurized columns in the split case, and $60,400 for the pump to bring the flow up to 4.2 atm before entering the pressurized column.

This results in an increased installed capital cost of $654,300 for the split beer column configuration with respect to the base case.

The steam re-compressors are taken to have a total installed cost of $300/kW of capacity. This cost represents the installed cost of the assembly, which includes the driving motor, compressor, and all necessary interfaces and accessories.

With these capital costs the simple payback period can be calculated for each of the evaluated scenarios. This payback period is the increased capital cost divided by the increased annual revenue. The resulting period indicates how many years of operation are required to recover the increased capital investment. Since these modifications are minor with respect to the total cost of the plant and are expected to be recovered in a short time period, the time value of money is not included in this calculation.

The plant is assumed to operate 8,400 hours per year and the assumed selling price for the electricity is 6.5 cents per kWhr. This electricity selling price approximates the rate at which utilities in Georgia are buying biomass derived electricity from independent producers. This rate would be expected to rise dramatically if either a renewable portfolio standard (RPS) or any carbon tax were applied at either a state or federal level. The RPS would dictate that electric utilities derive a stated percentage of their electricity from renewable sources. With the limited renewable resources in Georgia, biomass derived electricity would have great value.

It is shown in the following section that the pine to ethanol process studied here exhibits dramatic improvements over traditional methods with respect to net carbon emissions. Any form of a carbon tax would also make this biomass derived electricity more valuable.

These increases in the value of the excess electricity sold to the grid are not analyzed in this study, however electricity selling prices in the 10-15 cents per kWhr range would be a reasonable expectation if both legislations were enacted and would linearly decrease the payback periods calculated below. The simple payback periods for these configurations with the current approximate electricity selling price are shown below in Table 25.

Table 25: Payback periods for Alternate Scenarios

	Alternate Scenarios						
	2	**3**	**4**	**5**	**6**	**7**	**8**
Added Electricity Production (kW)	1,485	2,142	3,699	2,993	3,876	3,735	388
Operating Hours/yr	8400	8400	8400	8400	8400	8400	8400
Selling Price ($/kWhr)	0.065	0.065	0.065	0.065	0.065	0.065	0.065
Yearly increase in Revenue ($)	810,810	1,169,532	2,019,654	1,634,178	2,116,296	2,039,310	211,848
Increased Capital Cost ($)	654,300	1,488,900	3,383,700	1,261,200	2,335,200	2,681,100	5,717,400
Payoff Period (yrs)	**0.81**	**1.27**	**1.68**	**0.77**	**1.10**	**1.31**	**27.0**

As previously shown, the scenario which best reduces the plant energy demand is Scenario 6, as a result of its largest excess electricity available for sale. The above table shows that this scenario has a simple payback period of 1.10 years, showing that the increased capital cost of the split beer column and vapor re-compressors will be recovered in just over one year of operation.

Chapter 8: Net Energy and Carbon Life Cycle Assessments

8.1 Net Energy Analysis

As stated in the project scope, the two goals of this study are to increase the cost effectiveness and to improve the net energy ratio (NER) of pine ethanol. It was stated that both of these objectives can be achieved through the single parameter of increasing the amount of excess electricity sold as a byproduct. This has been performed in the previous section, and shown to increase the cost effectiveness of pine ethanol through the metric of the simple payback period.

The second goal of increasing the NER of pine ethanol can now be analyzed. The NER is the ratio of the life cycle energies of all of the outputs to all of the inputs. For this facility the primary input is pine chips, with secondary inputs being sulfur dioxide, yeast, and enzymes. The primary outputs are transportation grade ethanol and electricity, with secondary outputs being turpentine and furfural.

The NER does not simply ratio the heating values of the outputs and inputs, but evaluates the life cycle energies. The energy required per mass of pine chips is not simply how much heat the chips would give off if burned, but the amount of energy required to make the chips and transport them to the production facility. This life cycle energy includes the site preparation of the land, the planting of, managing of, harvesting of, chipping of, and transportation of the wood to the ethanol facility. This also includes the possibility of the fertilization of the land.

With the assistance of the Georgia Forestry Commission numbers were gathered which represent the agricultural inputs to the production of pine chips. The production of trees in Georgia are separated into three categories; natural stand, low intensity, and high intensity. Pine grown on a natural stand has no site preparation work, no tree planting, no fertilization, and only the merchantable wood is harvested. Low intensity pine has site preparation in the form of herbicides, hand tree planting, no fertilization, and only the merchantable wood is harvested. Finally, high intensity pine has mechanical site preparation, mechanical tree planting, fertilization, and both merchantable and non-merchantable wood is harvested. Table 26 below shows the energy requirement for pine grown under each set of conditions.

Table 26: Agricultural inputs to pine (gallons diesel/green ton pine) [5, 12, 17]

	Natural Stand	Low Intensity	High Intensity
Site Preparation	-	0.15	0.64
Tree Planting	-	-	0.42
Fertilizing	-	-	0.14
Stump to Truck transport	0.41	0.41	0.41
Chipping	0.41	0.41	0.41
Farm to Facility transport	0.58	0.58	0.58
Total	**1.40**	**1.55**	**2.60**

The above table shows the number of gallons of diesel fuel required to bring a green ton (~1/2 dry ton) of pine to the door of the ethanol facility. It is assumed that in the state of Georgia each of these three methods are used in equal proportion, and therefore the value for the amount of fuel used per green ton is taken as the average of the three (1.85 gallons diesel/green ton pine).

The secondary inputs to this process are sulfur dioxide, yeast, and enzymes. To obtain the life cycle energy required to produce and deliver these materials to the ethanol facility the economic

input output life cycle assessment (EIOLCA) tool was consulted[6]. This tool was developed and is managed by the Carnegie Mellon University Green Design Institute. It attempts to document information about the life cycle creation of products and services and includes the life cycle energy as well as the amount of green house gases emitted. The life cycle energy of these secondary inputs as shown in the EIOLCA tool is shown below in Table 27. The energy is shown with respect to the monetary value of the product and therefore must be converted.

Table 27: Life Cycle Energy of Secondary Inputs

	Energy (TJ/$MM)	Unit Cost ($/lbs)	Flow Rate (lbs/hr)	Specific Energy (BTU/gallon EtOH)
Sulfur Dioxide	36.1	0.115	98	58.0
Yeast	12.1	0.165	772	219.8
Enzymes	21.7	0.694	620	1,331.3

The final input to the process is the transportation of the ethanol product to the blending station. It is assumed that the vehicle carries 25 tons of ethanol per shipment, travels 5.5 miles per gallon of diesel, and travels 50 miles each direction, resulting in an energy charge of 2,341.7 BTU of diesel fuel per gallon of ethanol.

The primary outputs of this process are ethanol and electricity. The life cycle energy of the production of ethanol is what is being evaluated, and as a result for this evaluation its energy value is simply its lower heating value, which is approximately 76,330 BTU/gallon.

The life cycle energy of electricity is the primary energy that was required to originally produce the electricity. To obtain this value information was taken from the EIA (Energy Information Agency, US DoE)[7] as to how much primary fuel was used to produce electricity in a year, and

how much electricity was created during this year. The ratio of these two values gives the US grid average ratio of primary energy per electrical energy. The information obtained from the EIA 2005 annual report is shown below in Table 28. The heating value of the fuels was taken from the GREET 1.7 model produced by Argonne National Laboratory (ANL) [10].

Table 28: Primary fuel usage for 2005 US electricity production

Primary Fuel	2005 Amount	Unit	Heating Value	Units	Energy (KJ)
Coal	1,045,878	thousand tons	19,546,300	BTU/ton	2.156E+16
Petroleum	211,256	thousand barrels	129,670	BTU/Gallon	1.214E+15
Natural Gas	6,486,761	millions of cubic feet	983	BTU/SCF	6.726E+15
Other Gases	176,906	millions of BTU's	-	-	1.866E+11
Total					**2.950E+16**

This information shows that 2.950e16 KJ's of primary energy were used to produce electricity in the US in 2005, not including renewables and nuclear energy. The EIA also provides the value that 4,054,688 thousand MWhr's of electricity were produced in 2005, which is equal to 1.460e16 KJ of electrical energy. The ratio of these two values states that for every 2.021 units of fossil fuel combusted, 1 unit of electrical energy was sent into the electrical grid in 2005. This value does not include transmission losses, which is the appropriate value for this analysis because the excess electricity byproduct is measured as it enters the grid.

The life cycle energies of the secondary outputs of furfural and turpentine were calculated in the same manner as the secondary inputs. The information from the EIOLCA model and the corresponding specific energy per unit ethanol is shown below in Table 29.

Table 29: Life Cycle Energy of Secondary Outputs

	Energy (TJ/$MM)	Unit Cost ($/lbs)	Flow Rate (lbs/hr)	Specific Energy (BTU/gallon EtOH)
Furfural	22.7	0.50	366	592
Turpentine	3.78	0.139	144	10.8

The life cycle energy for all of the inputs and all of the outputs to this process have been calculated. The NER can now be taken as the sum of outputs divided by the sum of the inputs. The NER for the base case and each of the alternate scenarios is shown below in Table 30.

Table 30: Net Energy Ratio of Base Case and Alternate Scenarios (BTU/Gallon EtOH)

	Base	**Alternate Scenario**						
Inputs	**Case**	**2**	**3**	**4**	**5**	**6**	**7**	**8**
Site preparation, planting, harvesting, chipping, transporting to facility	5,957	5,957	5,957	5,957	5,957	5,957	5,957	5,957
Sulfur dioxide addition	58.0	58.0	58.0	58.0	58.0	58.0	58.0	58.0
Yeast addition	205.5	205.5	205.5	205.5	205.5	205.5	205.5	205.5
Enzyme addition	1,331	1,331	1,331	1,331	1,331	1,331	1,331	1,331
Transport to Blending Station	2,342	2,342	2,342	2,342	2,342	2,342	2,342	2,342
Total Energy Input	**9,894**	**9,894**	**9,894**	**9,894**	**9,894**	**9,894**	**9,894**	**9,894**
Outputs								
Ethanol	76,330	76,330	76,330	76,330	76,330	76,330	76,330	76,330
Furfural	592	592	592	592	592	592	592	592
Turpentine	10.8	10.8	10.8	10.8	10.8	10.8	10.8	10.8
Electricity	29,847	31,388	32,069	33,684	32,952	33,868	33,722	30,249
Total Energy Output	**106,780**	**108,321**	**109,002**	**110,617**	**109,885**	**110,801**	**110,655**	**107,183**
NER	**10.79**	**10.95**	**11.02**	**11.18**	**11.11**	**11.20**	**11.18**	**10.83**

It is shown that the NER has improved for every scenario with respect to the base case. It can be noted that every energy value is the same for each of the configurations, with the exception of the electricity output. This is because the rest of the plant is fixed during these scenarios,

resulting in the same ethanol yield as well as the same rate of input addition and output extraction. Since the electricity output is the only value which changes between the scenarios, it is of no surprise that Scenario 6 shows the highest NER, since it has the highest value of excess electricity sold as a byproduct.

For perspective, it has been shown by two independent sources that current corn ethanol facilities achieve a NER of approximately 1.29 to 1.65[11] and 1.1 to 1.5[8]. This is a result of the energy intensive farming of the corn as opposed to the pine, as well as the increased byproduct energy value of the electricity versus dried distillers grains with corn ethanol.

As another reference, the life cycle NER of gasoline is approximately 0.76[9]. This indicates that 24% of the energy content in the crude oil is consumed to remove the oil from the ground, transport it to a refinery, input energy to refine the crude to gasoline, and transport the gasoline to the filling station. This indicates that while the production of a liquid transportation fuel needs to have a NER of greater than 1.0 to produce net renewable fuel, it only has to have a NER greater than 0.76 to be an improvement over current gasoline production, from an energy viewpoint.

The NER's of the base case, alternate scenarios, gasoline, and corn ethanol including deviation among reviewed reports are shown graphically below in Figure 16.

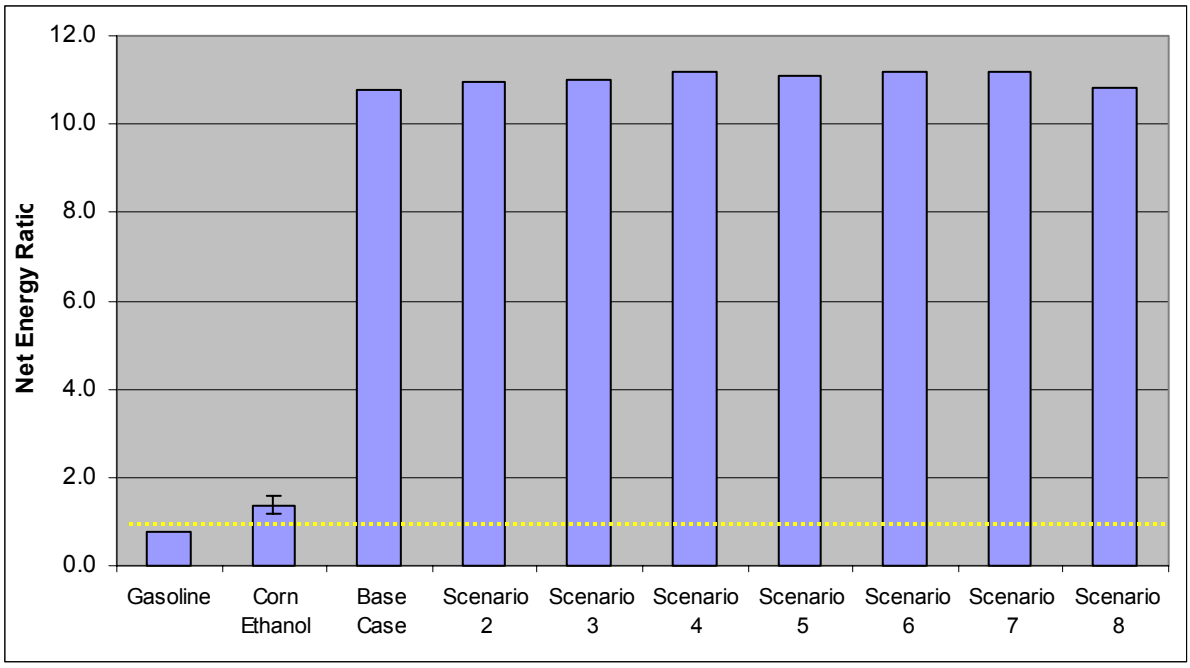

Figure 16: NER's of base case, alternate scenarios, and corn ethanol

The yellow line on the above figure represents the break even line, where the amount of non-renewable energy required for a process is equal to the renewable energy output from that process. Any process with a NER above 1.0 is said to have a net energy gain, with any process below resulting in a net energy loss. This analysis shows that pine ethanol, among any of the scenarios, has a NER an order of magnitude greater than corn ethanol, which in turn has a NER significantly higher than that of gasoline.

8.2 Net Carbon Analysis

The net carbon analysis is performed in much the same manner as the net energy analysis. However, instead of weighing the outputs against the inputs, the total life cycle emission of carbon is compared against the total life cycle emission of carbon by a comparable process

which creates the same outputs. In this process the outputs are a liquid transportation fuel and grid electricity. A comparable process to create a liquid transportation fuel is the production of gasoline, and a comparable process to the production of grid electricity is the US grid average electricity production.

Therefore, this analysis compares the release of carbon from the production and combustion of ethanol and the production of electricity in this facility versus the release of carbon from the production and combustion of gasoline and the production of grid electricity. To begin with, the life cycle carbon release of the secondary inputs and outputs is calculated. These emissions are calculated from the EIOLCA model as in the net energy analysis, and the values are shown below in Table 31.

Table 31: Life Cycle Energy of Secondary Inputs

	Carbon Dioxide Emissions (Tonnes/$MM)	Unit Cost ($/lbs)	Flow Rate (lbs/hr)	Specific Release (lbs CO2/gallon EtOH)
Sulfur Dioxide	1,970	0.115	98	0.007
Yeast	518	0.165	772	0.022
Enzymes	1,330	0.694	620	0.190
Furfural	1,350	0.50	366	0.082
Turpentine	249	0.139	144	0.002

In the same manner as in the net energy analysis, the EIA supplied information[7] is used to calculate the average amount of carbon dioxide released per unit of US grid electrical energy fed into the grid. The EIA states that 2,513,609 thousand metric tones of carbon dioxide were emitted by the generation of electricity in 2005, and that 4,054,688 thousand MWhr's were sent into the grid in 2005. This equals 1.367 lbs of carbon dioxide emitted per kWhr.

The final pieces of information needed are the amount of carbon dioxide released during the combustion of gasoline and diesel fuels. These are provided in the GREET 1.7[10] model developed by ANL as 26.3 lbs of carbon dioxide released per gallon of diesel combusted, and 23.0 lbs per gallon of gasoline[7]. These reflect the life cycle net releases of carbon dioxide for the two fuels, as opposed to the amount of carbon dioxide released simply during combustion.

The amount of carbon dioxide released by the production of ethanol and electricity, subtracted by the amount of carbon dioxide emissions avoided, are shown below in Table 32.

Table 32: Net CO2 Emissions of Base Case and Alternate Scenarios (lbs CO2/Gallon EtOH)

	Base	Alternate Scenario						
CO2 Emissions Released	**Case**	**2**	**3**	**4**	**5**	**6**	**7**	**8**
Site preparation, planting, harvesting, chipping, transporting to facility	1.220	1.220	1.220	1.220	1.220	1.220	1.220	1.220
Sulfur dioxide addition	0.007	0.007	0.007	0.007	0.007	0.007	0.007	0.007
Yeast addition	0.022	0.022	0.022	0.022	0.022	0.022	0.022	0.022
Enzyme addition	0.190	0.190	0.190	0.190	0.190	0.190	0.190	0.190
Transport to Blending Station	0.479	0.479	0.479	0.479	0.479	0.479	0.479	0.479
Net CO2 Released	**1.918**	**1.918**	**1.918**	**1.918**	**1.918**	**1.918**	**1.918**	**1.918**

It is noted that no carbon dioxide is emitted from the combustion of ethanol or from the flue gas from the electricity generation. This is because the carbon within these two streams comes solely from the carbon within the pine chips, which comes solely from the carbon dioxide that was pulled out of the atmosphere from the trees. The production of furfural and turpentine acts as avoided carbon dioxide emissions, meaning that because these chemicals are produced in this process, with carbon emissions already accounted for, they are not produced during traditional methods. The final note is that the net amount of carbon dioxide released is identical for each of

the scenarios and for the base case. This is because this emission is independent of the excess electricity sale, which is the only parameter which separates these scenarios in this analysis.

As stated, the purpose of this analysis is to compare two pathways for the production of a liquid transportation fuel and electricity on the basis of life cycle carbon emissions. The production and combustion of gasoline provides the same end use as the production and combustion of ethanol. However, these two fuels do not provide the same amount of energy for a given volume, therefore the fuels are scaled by this ratio and one gallon of ethanol is assumed to provide the same end use as 0.658 gallons of gasoline. This is conservative since it is only scaling the heating values of the fuels. Ethanol also has beneficial oxygenate and improved octane characteristics, both of which increase its combustion efficiency, however are not quantified in this analysis.

The comparison of the emission of carbon for the two parallel pathways is shown below in Table 33. The amount of carbon dioxide released during the production of pine ethanol and its corresponding electricity is shown as 'Net CO2 Released.' The amount of carbon dioxide that would be emitted to produce the same amount of electricity is shown as 'CO2 released through traditional electricity generation,' and the amount of carbon dioxide that would be emitted by producing and combusting enough gasoline to produce the same amount of heat as in a gallon of ethanol is shown as 'CO2 released through gasoline production and combustion.'

Table 33: Net CO2 Analysis (lbs CO2/Gallon EtOH)

	Base	Alternate Scenario						
	Case	2	3	4	5	6	7	8
Net CO2 Released	**1.92**	**1.92**	**1.92**	**1.92**	**1.92**	**1.92**	**1.92**	**1.92**
CO2 released through traditional furfural production	0.082	0.082	0.082	0.082	0.082	0.082	0.082	0.082
CO2 released through traditional turpentine production	0.002	0.002	0.002	0.002	0.002	0.002	0.002	0.002
CO2 released through traditional electricity generation	5.92	6.22	6.36	6.68	6.53	6.71	6.68	5.99
CO2 released through gasoline production and combustion	15.1	15.1	15.1	15.1	15.1	15.1	15.1	15.1
Total CO2 emissions avoided	**21.1**	**21.4**	**21.6**	**21.9**	**21.7**	**21.9**	**21.9**	**21.2**
Percent Reduction	**90.9%**	**91.0%**	**91.1%**	**91.2%**	**91.2%**	**91.2%**	**91.2%**	**91.0%**

The data in the above table shows that 90.9% of the carbon dioxide normally emitted by the production of gasoline and electricity is avoided by the production of ethanol and its associated electricity in the base case. This number varies throughout the alternate distillation scenarios, with a maximum of 91.2% in Scenario 6. This is not a large increase with respect to the base case, however understandable since with the ethanol yield fixed among the scenarios the only difference is the increase in the excess electricity available for sale.

As a comparison the amount of carbon dioxide emissions avoided through the production of ethanol from corn is approximately 28.7%, with values ranging from 4.0 to 45.0%, as calculated by Berkeley[8]. This value is the average carbon dioxide reduction calculated from a review of several studies on the production of corn ethanol which include the co-product credit of producing dried distillers grains (DDG's). The carbon dioxide reductions of the base case, the

alternate scenarios, and of the average case corn ethanol process are shown graphically below in Figure 17.

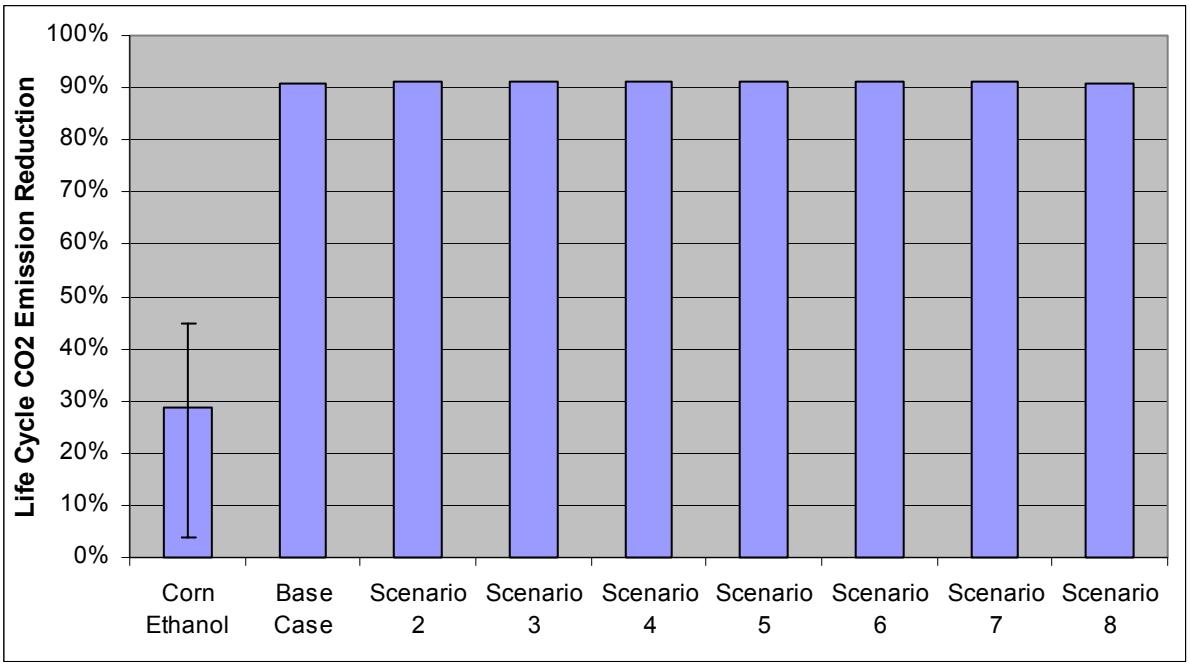

Figure 17: Bar graph of carbon emissions reductions in corn ethanol, base case, and alternate scenario processes

This analysis demonstrates that the pine ethanol process decreases the net carbon released to the atmosphere by over 90% when compared to the production of an equal energy value of gasoline and electricity.

It is noted that almost the entirety of the net carbon released in the pine ethanol process is from the production and combustion of diesel fuel. This fuel is used to power the transport vehicles as well as equipment used in the managing of the pine forests. These vehicles are powered by diesel fuel because of its current availability and the limited diversity of fuel available on such equipment. However, there is no reason that these vehicles can not be powered by ethanol, or

another renewable fuel. If this were the case the net carbon released during the production of the pine ethanol would drop to near zero, creating a 100% decrease in the net carbon emissions. The ethanol used by the vehicles would need to be subtracted from the total amount produced, but this would not change the fact that no new carbon would be added to the atmosphere.

The same applies for the NER analysis. If the transportation and forest managing equipment were fueled by ethanol the non-renewable energy input to the process would drop to near zero, creating an infinite NER, indicating that renewable fuel is created without any need for non-renewable input.

This situation could only apply when ethanol, or other renewable fuels, provides a large portion of the transportation fuel supply and is readily available throughout the US.

Chapter 9: Conclusions

Two main obstacles impeding the further growth of the ethanol industry have been addressed in this study; the cost-competitiveness with traditional fuels and the net energy ratio (NER). Through a series of alternate distillation and solids recovery scenarios the facility processing energy has been decreased in a manner as to increase the revenue from the byproduct sale of the excess electricity by over $2.1 MM/year. This increased revenue is achieved through a one time capital investment of $2.3 MM, resulting in a simple payback period of 1.1 years.

The NER of the process has been increased from a base case level of 10.8 to a value of 11.2. The level of both of these NER's is an order of magnitude greater than that of current corn ethanol plants and even of gasoline, ~1.3 and 0.8 respectively. As a result this pine ethanol process does not even enter in to the discussion of whether more non-renewable energy is consumed in the process than is returned as renewable energy. In conjunction with this increased NER, the amount of net carbon emissions reduction has been increased from 90.9% to 91.2% through the alternate distillation scenarios, compared with the current corn ethanol reduction of ~29%.

It is noted that in this analysis only the distillation and solids recovery section was modified. While this is the location of the vast majority of the energy demand, varying parameters in other sections can have a strong impact on the magnitude of this demand. One of the most influential parameters is the SSF solids loading, which controls the amount of dilution water added to the first SSF vessel. This analysis was performed with a fixed SSF solid loading of 10%, however if

this solids loading were decreased to 5% the total plant energy demand would increase by 92%, nearly doubling. If the solids loading were to increase to 14%, the total plant energy demand would drop by 25%. These large changes can make the improvements calculated in this analysis seem minor. However, it is likely that the trends of decreased plant energy demand experienced in this study are not exclusive to this level of solids loading, and can be applied to additional analyses at different solids loading levels. This would allow for the reductions in energy demand through the alternate distillation configurations and increased SSF solids loading to be applied concurrently.

These increases in byproduct electricity revenue and NER can increase the viability of the cellulosic ethanol industry and quicken its entry into the commercial market as it increases its competitiveness with traditional energy sources. This trend towards cost competitiveness also grows as traditional transportation sources, gasoline and diesel from petroleum, as well as traditional electricity generation sources, primarily coal and natural gas, continue to increase in price and face a growing resistance due to their contributions to climate change.

These two trends can help to ensure that this recent research into biofuels will not be a short-lived burst as with previous efforts, but instead will carry this field to an emergence as a self-sufficient competitive established industry.

Chapter 10: Recommendations

Additional analyses can be performed to confirm the findings discussed in this study, as well as to expand the breadth of this study beyond the feedstock of Southern Pine. To increase the applicability of this study it is recommended that the following future work be performed.

- Perform a second law of thermodynamics, or exergy, analysis on the entire process to confirm the findings in the comparison of the energy demands of the base case and the alternate scenarios
    - Additional heat integration configurations may become apparent under such an analysis
- Adapt the model developed in this study to expand the feedstock to cellulosic sources other than southern pine. Two sources to consider are corn stover and switchgrass.
- Perform this analysis at various SSF solids loadings to quantify the trade-off of increased fermentation yields with increased distillation energy demand.

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